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PROCESS AND ENVIRONMENTAL TECHNOLOGY FOR PRODUCING SNG AND LIQUID FUELS

Milton R. Beychok

PREPARED FOR
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By

Milton R. Beychok

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ABSTRACT

This report presents the process technology and environmental factors involved in the emerging industries for providing new supplemental energy supplies from non-conventional sources. It includes: (1) the production of substitute natural gas (SNG) from coal, crude oil and naphtha, (2) importing overseas gas supplies in the form of liquefied natural gas (LNG) and as liquid methanol, (3) the regasification of LNG, (4) the production of liquid fuels from oil shale, and (5) the liquefaction of coal to produce clean fuels. The report also includes introductory chapters to familiarize the reader with the technology of oil and gas processing, heat balances, fuel combustion and stack gases, thermal efficiencies, and water balances.

In most cases, the report identifies and quantifies the environmental emissions and effluents from each technology by specific examples from actual designs. It also presents a brief description of the process technology involved. Each of the sections on the individual technologies includes a recommended list of additional reading.

The report is oriented more towards providing environmental data rather than providing detailed process design factors. Insofar as possible, the report was written for a broad range of general readers instead of experienced process engineers.

The data provided is inter-disciplinary in nature since it covers air emissions, water disposition and effluent treatment, thermal balances, noise factors, solid wastes, and some of the socio-economic factors.

This report was written and submitted in fulfillment of Contract Number 68-03-2136, by Milton R. Beychok, Consulting Engineer, under the sponsorship of the Environmental Protection Agency. Work was completed as of December 1974.

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SECTION I

BACKGROUND

Based on current trends in the supply and demand of energy, the world faces a very serious shortage of petroleum crude oil. Middle Eastern politics have compounded the oil shortage for the industrialized nations of the world. But regardless of the ultimate resolution of the Middle East situation, the shortage between supply and demand will still exist and the energy 'crisis' will be with us for a long time to come. Unless we develop alternative energy supplies, we face a tremendous economic drain of our capital resources from the high cost of imported crude oil.

In the United States, the shortage of domestic crude oil is further compounded by an equally serious dwindling of natural gas supplies. Although we could supply most of our energy needs for hundreds of years by utilizing our vast reserves of coal, public concern with the environmental effects of coal mining and coal burning has worked against the direct use of coal for heat and power generation.

As a result of this situation, the U.S. government has initiated Project Independence to develop supplemental domestic energy sources and supplies. The nation's energy-supplying industries have undertaken a dramatic acceleration in the search for alternative methods of producing clean-burning gas and liquid fuels from a wide variety of domestic resources such as coal, oil shale and even municipal refuse. At the same time, these industries have intensified their search for overseas gas supplies which can be imported into the United States.

The purpose of this report is to explain in relatively simple terms the processes and technology currently available for: (1) producing substitute natural gas (SNG), (2) importing overseas gas supplies in the form of liquefied natural gas (LNG) or as liquid methanol, (3) the conversion of coal into low-sulfur oil and (4) the production of low-sulfur oil from oil shale. In the next few years, Federal and

State environmental agencies will undoubtedly become involved in regulating the air emissions, water effluents and solid wastes from these new 'emerging industries'.

This report is intended for use as a reference 'primer' by those environmental agency personnel who will be concerned with the environmental regulations for these new industries producing and transporting alternative energy supplies. Therefore, the report is oriented more towards the environmental factors of the various process technologies than towards detailed technical design factors. The report was not written for the engineer who is already expert in the technologies involved, but rather for those who need a broad overview and a generalized understanding in order to better evaluate the environmental factors involved in these new industries.

SECTION II

INTRODUCTION

Some knowledge of the composition of natural gas and crude oil is required to become familiar with the technology of producing alternative energy supplies. An understanding of the terminology, or technical 'jargon' used in the oil and gas industries is also needed. This section is intended to provide that basic introductory knowledge with a narrative discussion, followed by a glossary of terms and definitions.

> The traditional English engineering units of measurements and special 'trade' units are deeply ingrained into the terminology of the oil and gas industries. Those traditional units are retained throughout the narrative text of this report so that the reader may learn to understand the language of the oil and gas industries. However, all of the tables in the report are separately presented in English units and in the SI metric system (for example, Table 1E is in English units and Table 1M presents the same data in the SI metric system). In addition, a conversion table is included at the end of this section so that the reader may convert terms and units appearing in the text into metric units if desired.

HYDROCARBONS AND WHAT THEY ARE

Both raw natural gas and petroleum crude oil are mixtures of 'hydrocarbons', which are chemical molecules composed of hydrogen and carbon atoms. These molecules may be quite simple or quite complex in their structural arrangement (or 'linkage') of atoms. The carbon atoms may be linked together in short or long straight chains, or they may be linked in complex rings or cyclic arrangements. The usual simple hydrocarbons found in natural gas are called 'normal paraffins' or 'saturated hydrocarbons' because each carbon atom is linked to a maximum number of hydrogen atoms in accordance with the generic formula of C_nH_{2n+2}. For example, a saturated hydrocarbon with 3 carbon atoms would contain 8 hydrogen atoms.

Hydrocarbons with only a few carbon atoms (from 1 to 6) are referred to as 'light' hydrocarbons and those with more carbon atoms are 'heavy' hydrocarbons. The very light hydrocarbons (1 to 4 carbon atoms) are gases under natural conditions, although they can be liquefied by refrigeration or by a combination of compression and cooling. Each hydrocarbon has a specific formula and chemical name, but in common usage they are also frequently referred to by the number of carbon atoms they contain. The following list of hydrocarbons and their symbols is not by any means complete, because there are many molecular structural arrangements other than the normal, or saturated, hydrocarbons:

Chemical Name	Carbon Atoms Per Molecule	Common Usage Symbol
Methane	1	C_1
Ethane	2	C_2
Propane	3	C_3
Butane	4	C_4
Pentane	5	C_5
Hexane	6	C_6
Heptane	7	C_7
Octane	8	C_8

NATURAL GAS

Raw natural gas, as it occurs in nature, is predominantly methane but it almost always contains some of the other very light hydrocarbons (ethane, propane, and butane) and it may also contain some pentane, hexane and even heavier hydrocarbons. A raw natural gas which contains significant amounts of C_3, C_4, C_5, C_6 and heavier is called a 'wet gas', and one which contains mostly methane and some ethane (C_1 and some C_2) is called a 'dry gas'. The terms 'wet' and 'dry' denote in a very relative manner whether or not the gas contains hydrocarbons heavier than ethane.

The C_3 and C_4 (propane and butane) in a raw natural gas can be removed and recovered as liquids by relatively simple processing.

Those liquids are called LPG (liquefied petroleum gas) and are subcategorized as C_3 LPG and C_4 LPG, or Propane LPG and Butane LPG. When contained under pressure, LPG can be stored and shipped as a liquid for end-use as fuel.

The C_5, C_6, C_7 and heavier molecules in a raw natural gas can be very readily removed and condensed into a liquid commonly referred to as 'condensate' or as 'natural gasoline' because it can be used as a component of gasoline.

The C_2 (ethane) in a raw natural gas can also be removed and recovered, but it requires relatively complex processing. Whereas LPG and condensate are almost always removed and recovered from raw natural gas, the removal and recovery of ethane depends on how much is present and the economic viability of that recovery. Among other uses, ethane is used as a petrochemical feedstock in producing polyethylene plastics.

Raw natural gas usually contains other non-hydrocarbon gases, such as water vapor and the so-called 'acid-gases' which are hydrogen sulfide (H_2S) and carbon dioxide (CO_2). A natural gas with a relatively high H_2S content is called a 'sour' gas, while one with a low H_2S level is called a 'sweet' gas. In order to meet pipeline and end-use specifications, a raw natural gas must be processed, or 'treated', for removal of H_2S and CO_2 down to very low levels. LPG and condensate are usually removed and recovered because it is economically advantageous to do so, and because the end-use natural gas specification on heating value may require their removal. Water vapor must also be removed to meet pipeline gas specifications.

Treated gas specifications will vary from one project to another. However, as a broad generality, treated gas specifications will fall within the following ranges:

Methane	85 - 99+ vol %
Ethane and heavier	0 - 15 vol %
Inert gases	0 - 3 vol % *
H_2S	0.20 - 0.50 grains/100 SCF **
CO_2	0 - 2 vol %
Water vapor	As required by dewpoint specifications
Heating value	950 - 1150 BTU/SCF (HHV)

We can thus conclude that raw natural gas is not precisely definable in terms of composition and that it may contain a range of hydrocarbons, inert or non-combustible gases, acid gases and water vapor. Before being shipped to the end-use market, the raw gas is usually processed, treated and dehydrated for: (1) recovery of LPG, condensate (natural gasoline), and perhaps for ethane and (2) removal of the acid gases (H_2S and CO_2) and water vapor. The plants which process the raw natural gas into a treated end-use gas are variously called 'field gas plants', 'gas treating' plants or 'gas recovery' plants. The liquids recovered include C_3 LPG, C_4 LPG and condensate. Sometimes all of these are collectively called NGL (natural gas liquids). When the processed or partially processed natural gas is liquefied for overseas shipment it is called LNG (liquefied natural gas).

After all that processing, it is really a contradiction in terms to call the end-use product a 'natural' gas -- about all we can say is that it is predominantly methane.

CRUDE OIL

Crude oil as pumped from underground wells is a very broad mixture of hydrocarbons, ranging from about 5 carbon atom molecules to

---

* Typical inert gases that may be present are nitrogen, helium, argon, etc. Occasionally, some oxygen may be present.

** 0.00068 - 0.00169 wt % H_2S.

20 or more carbon atom molecules (C_5 to C_{20+}).

Both natural gas and crude oil occurring in nature may be found alone or in combination. A raw natural gas may contain a good bit of C_5, C_6, and C_7 as well as methane, ethane, propane and butane; on the other hand, raw crude oil may contain some 'dry' gas (C_1 and C_2) as well as C_3 and C_4. When the oil well produces large amounts of raw natural gas as well as crude oil, the two are usually separated at or near the well-head, and the separated raw natural gas is called 'associated gas' (as distinguished from non-associated gas produced from a gas well). When the crude oil contains only nominal amounts of light gas, it is generally shipped "as is" to an oil refining plant, where the gas is ultimately separated from the oil and used as refinery fuel.

Since crude oil is such a broad hydrocarbon mixture, and since the light hydrocarbons have lower boiling points than the heavy hydrocarbons (see Table 1), the first step in refining crude oil is to simply 'distill' or boil off the light hydrocarbons. The light gases (C_1 through C_4) distill off first since they have the lowest boiling points. They are processed in a gas recovery and treating plant within the refinery to produce fuel gas (C_1 and C_2) and LPG (C_3 and C_4). The next material which boils off is naphtha or gasoline, a mixture of liquid hydrocarbons boiling over the range of 100^oF to about 400^oF and containing hydrocarbons ranging from C_5 to C_{11} (see Table 1). The next hydrocarbon mixture to boil off is diesel oil containing C_{10} to C_{15} and boiling over the range of 350^oF to 550^oF. Following that, a light fuel oil, or 'distillate oil', is boiled off.

At this point, the remaining residual crude oil would begin to suffer degradation to coke if the boiling were continued at too high a temperature. So the residual crude oil from the 'atmospheric distillation' unit is next distilled under vacuum to produce heavy fuel oil. The light and heavy fuel oils contain hydrocarbons in

TABLE 1B - FUEL HEATING VALUES - ENGLISH UNITS

Hydrocarbon	Carbon Atoms Per Molecule	Weight % Carbon	Weight % Hydrogen	Heating Value Net Btu/lb	Heating Value Gross Btu/lb	Heating Value Gross Btu/SCF (1)	Boiling Point @ Atm. Pressure °F
Methane	1	75	25	21,500	23,900	1010	-259
Ethane	2	80	20	20,400	22,300	1765	-128
Propane	3	82	18	19,900	21,700	2520	- 44
Butane	4	83	17	19,700	21,300	3260	31
Pentane	5	83.4	16.6	19,500	21,100	--	97
Naphtha	5 to 11	85	15	19,200	20,700	--	100 to 400 (2)
Diesel Oil	10 to 15	87	13	18,300	19,500	--	350 to 550 (3)
Fuel Oil	16 plus	88	11(5)	17,600	18,700	--	650 plus (4)
Others							
Hydrogen	--	0	100	--	61,100	320	-423
Petroleum Coke	--	96	2(6)	--	15,200	--	--
Coal (New Mex.)							
As mined	--	51	4(7)	--	8,800	--	--
Dry & Ash-free		77	6(8)	--	13,200	--	--

(1) Standard cubic feet (SCF) of gas, measured at 60°F and atmospheric pressure
(2) Initiates boiling at 100°F and ends boiling at 400°F
(3) Initiates boiling at 350°F and ends boiling at 550°F
(4) Initiates boiling at about 650°F and ends boiling much higher
(5) The remaining 1% is primarily sulfur and nitrogen, with traces of heavy metals
(6) The remaining 2% is primarily sulfur, nitrogen and ash, with traces of heavy metals
(7) The remaining 45% is mostly moisture, ash, sulfur, nitrogen, bound oxygen and traces of many other elements
(8) The remaining 17% is mostly sulfur, nitrogen, bound oxygen and traces of other elements

TABLE 1M - FUEL HEATING VALUES - METRIC UNITS

Hydrocarbon	Carbon Atoms Per Molecule	Weight %		Heating Value			Boiling Point @ Atm. Pressure °C
		Carbon	Hydrogen	Net Kcal/kg	Gross Kcal/kg	Gross Kcal/nm^3 (1)	
Methane	1	75	25	11,955	13,290	9,500	-162
Ethane	2	80	20	11,340	12,400	16,600	- 89
Propane	3	82	18	11,065	12,065	23,700	- 42
Butane	4	83	17	10,955	11,845	30,665	- 0.6
Pentane	5	83.4	16.6	10,840	11,730	--	36
Naphtha	5 to 11	85	15	10,675	11,510	--	38 to 204 (2)
Diesel Oil	10 to 15	87	13	10,175	10,840	--	177 to 288 (3)
Fuel Oil	16 plus	88	11(5)	9,785	10,395	--	343 plus (4)
Others							
Hydrogen	--	0	100	--	33,970	3,010	-217
Petroleum Coke	--	96	2(6)	--	8,450	--	--
Coal (New Mex.)							
As mined	--	51	4(7)	--	4,890	--	--
Dry & Ash-free	--	77	6(8)	--	7,340	--	--

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(1) Normal cubic metre (nm^3) of gas, measured at 0°C and atmospheric pressure
(2) Initiates boiling at 38°C and ends boiling at 204°C
(3) Initiates boiling at 177°C and ends boiling at 288°C
(4) Initiates boiling at about 343°C and ends boiling much higher
(5) The remaining 1% is primarily sulfur and nitrogen, with traces of heavy metals
(6) The remaining 2% is primarily sulfur, nitrogen and ash, with traces of heavy metals
(7) The remaining 45% is mostly moisture, ash, sulfur, nitrogen, bound oxygen and traces of many other elements
(8) The remaining 17% is mostly sulfur, nitrogen, bound oxygen and traces of other elements

the range of C_{20} and heavier and boil at temperatures of 650°F and higher (see Table 1).

All of the materials boiled off in the atmospheric distillation unit (or 'crude unit') and in the vacuum unit are referred to as 'virgin' -- virgin naphtha, virgin diesel, virgin fuel oils, etc. The final residual material is called 'vacuum residuum' or 'vacuum bottoms' or simply 'resid'.

Almost all of the virgin materials are then processed through a host of other steps, which can be combined into hundreds of different 'refinery configurations' to produce any desired 'slate' or end-products. The typical product slate for a 'motor fuels refinery' may include:

1. Two or three grades of gasoline (regular, premium, low lead, etc.)
2. C_3 and C_4 LPG
3. Diesel oil and/or jet fuel, which are quite similar
4. A range of industrial heating fuel oils (No. 2 distillate oil, No. 6 fuel oil, etc.)
5. By-product sulfur
6. Asphalt, tar and/or petroleum coke

Other slates would be produced by 'lube oil refineries', 'SNG refineries', 'fuel oil refineries', etc. It is beyond the scope of this basic handbook to attempt an explanation of all the available refining processes or of the numerous possible combinations of those processes into different refinery configurations.

Returning to gas wells and oil wells: most of the naturally occurring C_1 and C_2 are obtained from gas wells, while most of the naphtha and heavier distillate oils are derived from crude oil wells. C_3 and C_4 LPG may come from either gas or oil wells. Light naphtha (C_5, C_6 and C_7) may also come from either source, but by far the largest part of the total naphtha production comes from crude oil. So, if there is any typical breakpoint between the composition of raw natural gas and raw crude oil, it is between butane (C_4) and pentane (C_5). However, as mentioned earlier, either raw natural gas or raw crude oil may contain hydrocarbons normally found in the other.

Summarizing some of the commonly used terms mentioned in the preceding discussion:

Symbol	Physical Form	Common Usage Names
C_1	gas	Methane (primary constituent of natural gas)
C_1	liquid	Liquefied natural gas (LNG)
C_1 and C_2	gases	Dry gas
C_3 and C_4	liquids	Liquefied petroleum gas (LPG)
C_1, C_2, C_3, C_4	gases	Light hydrocarbons, or 'light ends' (in a refinery)
C_5 - C_8	liquid	Condensate or natural gasoline (when derived from raw natural gas)
C_5 - C_8	liquid	Light naphtha or light gasoline (when derived from crude oil)
C_3, C_4, C_5-C_8	liquids	Natural gas liquids (NGL) as a collective term (when derived from natural gas)
C_5 - C_{11}	liquid	Naphtha or gasoline (or full-range naphtha and full-range gasoline)
C_{10} - C_{15}	liquid	Diesel oil, jet fuel, light distillate oil, light furnace oil
C_{16+} - C_{20+}	liquid	Fuel oil, heavy fuel oil, heavy distillate oil, heavy furnace oil, bunker oil, resid
H_2S, CO_2	gases	Acid gases

SUBSTITUTE NATURAL GAS

Substitute natural gas (SNG) is a methane-rich gas manufactured from coal, LPG, naphtha, crude oil or other materials containing carbon.

The carbon in the feedstock material is chemically combined with hydrogen to yield methane (CH_4). The hydrogen is usually derived from steam and may be augmented by hydrogen already present in the feedstock (as in the case of LPG, naphtha or other hydrocarbons). In other words, the chemical addition of hydrogen to a heavy hydrocarbon changes it to a lighter hydrocarbon and, if sufficient hydrogen is added, we obtain methane -- the lightest of the hydrocarbons.

Various names have been used to describe manufactured methane, such as:

Substitute natural gas (SNG)
Synthetic natural gas (SNG)
Substitute pipeline gas (SPG)
Synthetic pipeline gas (SPG)
Supplemental pipeline gas (SPG)

In any event, they are all meant to describe a manufactured gas containing about 97% methane, having a higher heating value (HHV) of about 1000 Btu/SCF, and meeting the same end-use specifications as pipelined natural gas.

In the United States, gas distribution systems and gas-burning appliances have been designed to handle natural gas which has an HHV of about 1000 Btu/SCF (a gas of about that heating value is referred to as 'high Btu gas') and a plant producing SNG for pipeline distribution must therefore manufacture a gas meeting that specification.

If the gas were to be used at the point of manufacture, as a fuel source in electrical power generation for example, then it could be a 'low-Btu gas' with an HHV of about 350 Btu/SCF or even less. A low-Btu gas may contain varying amounts of carbon monoxide, carbon dioxide and hydrogen, as well as varying amounts of methane. Both carbon monoxide and hydrogen have HHVs of about 320 Btu/SCF and carbon dioxide has no heating value at all, so varying the amounts of the different components can produce any heating value desired below that of methane.

It is obviously less costly to produce a low-Btu gas than a high-Btu gas, and this may make it attractive for some industries to use a manufactured low-Btu gas.

TABLE 2 -- ENGLISH-TO-METRIC UNIT CONVERSIONS

Multiply This	By This	To Obtain This	
lbs	0.4536	kg	kilograms
short tons	0.9072	metric tons	metric tons (1000 kg)
short tons	907.2	kg	kilograms
inches	2.54	cm	centimetres
feet	0.3048	m	metres
statute miles	1.609	km	kilometres
gallons	3.785	l	litres (1000 litres = 1 m^3)
barrels	0.1590	m^3	cubic metres
Btu	0.252	kcal	kilocalories
SCF	0.02679	nm^3	normal cubic metres
Btu/lb	0.5556	kcal/kg	kilocalories/kilogram
Btu/CF	8.899	kcal/m^3	kilocalories/cubic metre
Btu/SCF	9.406	kcal/nm^3	kilocalories/normal cubic metre
10^9 Btu/day	252	Gcal/day	gigacalories/day
10^6 Btu/day	252	Mcal/day	megacalories/day
MM Btu/hr	252	Mcal/hr	megacalories/hour
SCFD	0.02679	nm^3/day	normal cubic metres/day
MM SCFD	0.02679	10^6 nm^3/day (Mnm^3/day)	million normal cubic metres/day (mega normal cubic metres/day)
SCF/MM Btu	0.1063	nm^3/Gcal	normal cubic metres/gigacalorie
lbs/MM Btu	1.8	kg/Gcal	kilograms/gigacalorie
lbs/CF	16.02	kg/m^3	kilograms/cubic metre
psi	0.07031	kg/cm^2	kilograms/square centimetre
gpm	0.227	m^3/hr	cubic metres/hour
acre-ft/year	0.1408	m^3/hr	cubic metres/hour
horsepower	745.7	W	watts
nautical miles	1.852	km	kilometres
knot	1.852	km/hr	kilometres/hour

(a) A SCF of gas is measured at 60°F and atmospheric pressure, and a nm^3 of gas is measured at 0°C and atmospheric pressure.

(b)

Exponential	English	SI Metric
10^3	M or thousand	k or kilo
10^6	MM or million	M or mega
10^9	billion (U.S.)	G or giga
10^{12}	billion (U.K.)	T or tera

SECTION III

HEAT BALANCES, COMBUSTION GASES AND OVERALL WATER BALANCES

HEAT BALANCES

Most hydrocarbon processing plants require the extensive use of fuel-fired process heaters and furnaces to supply reaction and distillation heat. In general, they also require fuel-fired auxiliary steam generators to provide steam for reaction, distillation and heating, and for energy to drive compressors and pumps.

The intermediate process streams within a hydrocarbon processing plant often need cooling before entering a subsequent process step. Usually, intermediate and final products and byproducts must also be cooled before storage or shipment.

These demands for both heating and cooling within the process plant afford many opportunities for 'heat recovery' by exchange of heat between process streams. A well-designed plant will take every advantage of such opportunities to recover heat. For example, the crude oil entering a refinery atmospheric distillation tower (the 'crude unit') must be heated to about 700-750°F and, in a well-designed crude unit, about 60% of that heat will be obtained by exchange with the hot products distilled from the crude oil (this exchange also provides much of the cooling required for those hot products).

Many hydrocarbon processing plants will also recover heat from high-temperature process streams by heat exchanging those streams with water to generate steam; such a heat exchanger is usually called a 'waste heat boiler' (WHB).

Steam for driver energy, distillation and heating can often be utilized at various temperature and pressure levels. Again, this affords many opportunities to 'reuse' the steam so as to extract the maximum available energy from it. For example, most large refineries, petrochemical and gas treating plants will generate relatively high-pressure steam (600-900 psig) for use in turbines to drive large compressors and pumps. Whenever possible, these turbines will exhaust their steam at back-pressures of about 50-150 psig (rather than under vacuum, as is done in large power plants). The exhaust steam from these 'back-pressure turbines' is then reused as distillation and heating steam in the process units, wherein the latent heat of the steam is extracted by condensation.

By taking maximum advantage of the heat exchange opportunities, and by using back-pressure turbines to supply driver energy plus distillation steam, a well-designed oil refinery achieves a 90% overall thermal efficiency*. Contrast this with a large fuel-fired electric power generating plant, which exhausts steam from its turbo-generators under vacuum to derive as much turbine-driving energy as possible. The power plant has no place to utilize the vacuum exhaust steam and must condense it either with cooling water or air. The result is that the steam's latent heat (about 900 Btu lb of steam) is lost and this explains why a fuel-fired power plant has an overall thermal efficiency of only 32-42%.

Why this discussion of process heat recovery and steam reuse? Because it is important for those who undertake the environmental analysis of a process plant to understand that:

-- Both a detailed and an overall heat balance are needed. These balances provide the basis for an independent judgement of how well the plant design takes advantage of heat exchange opportunities.

-- A plant steam balance is also needed to provide an independent judgement of heat conservation within the plant design.

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* Defined as the percentage of the total input heating value (of the feedstocks plus the plant fuel) that is recovered as product and by-product heating value, or equivalent energy.

- -- The plant's overall thermal efficiency is a key environmental consideration when comparing the plant with its alternatives.
- -- The degree of process heat recovery affects the overall thermal efficiency and the overall demand for cooling water and boiler feedwater. The demand for fresh water supply is an important environmental consideration in itself and relates almost directly to the amount of plant effluent water to be handled and disposed.

HEATING VALUES

The analysis of a project's heat balance and its overall thermal efficiency requires knowledge of the heating values of the typical hydrocarbons involved in a project. Table 1 lists the HHVs (higher heating values) for the typical fuels, feedstocks and products that will be encountered. It is interesting to note that:

- -- As the hydrocarbons progress from light to heavy, both their hydrogen content and their Btu/lb heating values decrease. Obviously, heating value is a function of the carbon-to-hydrogen ratio.
- -- Although hydrogen is considered a 'low-Btu' gas on a volume basis (320 Btu/SCF), it is a very high-Btu fuel on the more meaningful weight basis (61,100 Btu/lb). The same inverse relationship between Btu/SCF and Btu/lb applies to the hydrocarbons as well. Lighter fuels have lower gas densities (i.e. less pounds per standard cubic foot) and hence their heating values are relatively higher on a weight basis than on a volume basis.
- -- The very low boiling temperature of the light C_1 to C_4 hydrocarbons explains why refrigeration (or compression and cooling) is required to liquify them in producing LNG and LPG. The boiling temperatures of the pentanes and heavier (100^{o}F and above) also explains why these are usually liquids.

The term 'higher heating value' (HHV) is synonymous with 'gross heating value' and represents the total heat released when a fuel is burned. The end products of fuel combustion include carbon dioxide (CO_2) formed from the carbon in the fuel, and water (H_2O) formed from the hydrogen. The water is vaporized by the heat release, using up some of the fuel's gross heating value and leaving a net heating value as the effective amount available for use. This net heating value is referred to as the 'lower heating value' (LHV). The difference between the higher and lower heating values of a fuel represents the heat 'lost' when the fuel is burned (by vaporization of the combustion product water) and which therefore cannot be recovered for heating use.

Quantitatively, the HHV is 5-10% higher than the LHV of a given fuel. It is not really necessary to discuss when or why it is more correct to use HHV or LHV, but it is pertinent to realize that heat balances or thermal efficiencies should be consistent in using one value or the other. The inconsistent use of heating values will result in significant errors.

COMBUSTION STACK GASES

When a fuel burns in air (which is essentially a mixture of oxygen and nitrogen), the carbon and hydrogen in the fuel combine with atmospheric oxygen to form carbon dioxide and water vapor. The combustion stack gases (or 'flue gas') are therefore composed mainly of carbon dioxide, water vapor and the residual atmospheric nitrogen. Since complete combustion requires some excess air, the stack gases also contain some oxygen and its additional co-mingled nitrogen. The stack gases will also contain some sulfur dioxide (SO_2) derived from sulfur in the fuel, some nitrogen oxides (NO_x) derived from nitrogen in the fuel and from atmospheric nitrogen, and some particulates derived from 'ash' in the fuel.

Determination of the amount of stack gas that results from the burning of fuel is needed in order to estimate SO_2, NO_x and particulate emission concentrations. Table 3 provides a quick and accurate estimate of the flue gas rates resulting from burning the indicated fuels and it is based on the fuel carbon-hydrogen ratios and heating values given in Table 1. In using Table 3, it should be noted that:

- -- The terms 'excess air' and 'O_2 in the flue gas' are not synonymous, as shown by their significant differences in the table. The difference between them is very often overlooked. (When an emission concentration regulation refers to 3% excess O_2, it probably means O_2 in the flue gas rather than the amount of excess air, but this should be carefully confirmed in each case).
- -- The amount of excess combustion air used in process heaters and steam generators will be in the range of 5-20%, while gas turbine combustors will use from 200-300%. Emission concentration values should always be corrected to the % excess air or to the % O_2 in the flue gas specified by the applicable emission regulations. Some emission regulations are expressed in lbs/MM Btu (or 10^6 Btu) which completely eliminates the problem of defining the flue gas basis. Ideally, all emission regulations should use lbs/MM Btu.
- -- Many emission concentration regulations fail to specify a 'wet' or 'dry' flue gas basis (i.e. one includes the combustion product water vapor, and one does not). Table 3 illustrates the significant difference between the two values.
- -- Coals from various geographic areas will vary widely in chemical composition and in heating value. Hence, the coal data in Table 3 are applicable only for the specified ultimate analysis and heating value.

OVERALL WATER BALANCES

An overall water balance is an accounting of the raw water supply to a project and the ultimate use or disposition of that water. An overall balance need not include the intermediate details of water flows within the plant, but such a water-flow diagram can be very

TABLE 3E -- FLUE GAS RATES - ENGLISH UNITS

		Wet Flue Gas (WFG)			Dry Flue Gas (DFG)		
	HHV, Btu/lb	Rate, SCF/MM Btu	vol %		Rate, SCF/MM Btu	vol %	
			O_2	CO_2		O_2	CO_2
Methane @ 5% excess air	23,900	10,890	0.9	9.1	8,910	1.1	11.1
@ 15% excess air	23,900	11,830	2.5	8.4	9,850	3.0	10.1
Propane @ 5% excess air	21,700	10,715	0.9	11.1	9,130	1.1	13.0
@ 15% excess air	21,700	11,660	2.6	10.2	10,075	3.0	11.8
Naphtha @ 5% excess air	20,700	10,600	0.9	12.2	9,225	1.1	14.1
@ 15% excess air	20,700	11,540	2.6	11.2	10,170	2.9	12.8
Fuel Oil @ 5% excess air	18,700	10,795	0.9	13.8	9,690	1.1	15.3
@ 15% excess air	18,700	11,770	2.6	12.6	10,655	2.9	13.9
Coal(1) @ 5% excess air	8,800	11,590	0.9	15.8	10,440	1.0	17.5
@ 15% excess air	8,800	12,615	2.6	14.5	11,465	2.8	16.0

---

(1) As mined, and based on this ultimate analysis:

C = 51 wt%
H = 4 wt%
O = 9 wt%
H_2O = 12 wt%
Ash = 23 wt%
S = 1 wt%
100 wt%

TABLE 3M -- FLUE GAS RATES-METRIC UNITS

			Wet Flue Gas (WFG)			Dry Flue Gas (DFG)		
				vol %			vol %	
		HHV kcal/kg	Rate, $nm^3/Gcal^*$	O_2	CO_2	Rate, $nm^3/Gcal^*$	O_2	CO_2
Methane	@ 5% excess air	13,290	1,158	0.9	9.1	947	1.1	11.1
	@ 15% excess air	13,290	1,258	2.5	8.4	1,047	3.0	10.1
Propane	@ 5% excess air	12,065	1,139	0.9	11.1	971	1.1	13.0
	@ 15% excess air	12,065	1,239	2.6	10.2	1,071	3.0	11.8
Naphtha	@ 5% excess air	11,510	1,127	0.9	12.2	981	1.1	14.1
	@ 15% excess air	11,510	1,227	2.6	11.2	1,081	2.9	12.8
Fuel Oil	@ 5% excess air	10,395	1,148	0.9	13.8	1,030	1.1	15.3
	@ 15% excess air	10,395	1,251	2.6	12.6	1,133	2.9	13.9
Coal(1)	@ 5% excess air	4,890	1,232	0.9	15.8	1,110	1.0	17.5
	@ 15% excess air	4,890	1,341	2.6	14.5	1,219	2.8	16.0

---

(1) As mined, and based on this ultimate analysis:

C = 51 wt %
H = 4 wt %
O = 9 wt %
H_2O = 12 wt %
Ash = 23 wt %
S = 1 wt %
100 wt %

* Gcal = 10^9 calories

useful in lending credibility to the overall balance.

In most hydrocarbon processing plants, the overall water balance discloses that very little of the supply water is "destroyed" or disposed of irretrievably. In fact, most of the water is evaporated and returned to the regional atmosphere to eventually return as rainwater. Even water used to supply hydrogen (derived from steam) for producing methane SNG will return to the atmosphere as combustion product water vapor wherever that methane is burned.

For example, consider this water balance for a coal gasification plant design:

	gpm	%
Process consumption (supplying hydrogen)	520	10.2
Return to atmosphere via:		
Evaporation	3,310	64.9
Scrubbers and steam vents	240	4.7
	3,550	69.6
Disposal in mine reclamation	430	8.4
Other uses	600	11.8
	5,100	100.0

In this case, about 80% of the total water supply will return to the atmosphere from evaporation, venting and combustion of the product SNG. Also, this project was designed to have no discharge of wastewater.

As another example, this is the overall water balance for a project producing SNG from naphtha (with the total cooling needs provided by air-cooling):

	gpm	%
Process conversion (supplying hydrogen)	300	78.9
Vented with CO_2	17	4.5
Demineralizer and boiler blowdowns	50	13.2
Treated wastewater discharge	12	3.2
On by-product sulfur	1	0.2
	380	100.0

Here, about 83% of the water supply will return to the atmosphere via venting and from the burning of the SNG.

As exemplified in the two cases above, an overall water balance helps to put the project in proper perspective. As noted earlier, very little of the water supplied to a hydrocarbon processing plant is destroyed, since most of it eventually returns to the atmosphere as water vapor. The key environmental consideration is not how much water 'passes through', but rather how much is discharged as blowdown and wastewater, and what contaminants those discharges will contain.

SECTION IV

SNG FROM LPG AND/OR NAPHTHA

Substitute natural gas (SNG) can be produced from light hydrocarbons such as LPG, condensate or naphtha. Currently, there are at least three process technologies available for such plants: Lurgi's Gasynthan process; the British Gas Council's CRG process; and the Japanese MRG process. These processes can have an overall thermal efficiency as high as 97-98% (see Table 4). The three processes are quite similar, differing mainly in proprietary details and catalysts. A fairly typical design, using the Lurgi process, will be described in this section as illustrative of the technology.

The Lurgi process is quite simple. In brief, the end-product methane SNG is produced by a series of chemical reactions which combine the feedstock hydrocarbon with hydrogen to produce methane. Figure 1 typifies the reaction steps in a Lurgi design, when feeding naphtha:

-- Catalytic desulfurization (or 'hydrotreating') to remove sulfur from the naphtha by converting it to gaseous H_2S.
-- Gasification wherein naphtha, steam and hydrogen combine to form methane, carbon monoxide and carbon dioxide.
-- Methanation wherein carbon monoxide and hydrogen react to form additional methane.
-- Removal of CO_2 and water from the product SNG.
-- Sulfur recovery to convert gaseous H_2S into a salable by-product sulfur.
-- Hydrogen production from a portion of the crude SNG.
-- A boiler plant to generate steam, as well as other auxiliary utility services.

The plant as shown schematically in Figure 1, and described herein, feeds 20,700 barrels a day of feedstock and fuel naphtha to produce 100 MM SCFD of SNG with a heating value of 1000 Btu/SCF (HHV).

> This description of a naphtha SNG plant is based on a specific design and is not universally applicable. It is intended merely to illustrate the process involved.

TABLE 4E -- THERMAL EFFICIENCY OF SNG PRODUCTION FROM NAPHTHA - ENGLISH UNITS

OVERALL THERMAL EFFICIENCY

	FLOW RATE	10^6 Btu/Day
Naphtha Feedstock, lbs/hr	205,000	93,480(1)
, (barrels/day)	(18,950)	
Naphtha Fuel, lbs/hr	19,100	8,710(1)
, (barrels/day)	(1,765)	
TOTAL INPUT		102,190
SNG Product, SCFD	100×10^6	100,000(2)

Thermal Efficiency = (100,000/102,190)(100)
= 97.8%

FUEL REQUIREMENTS

	10^6 Btu/hr HEAT RELEASE	lbs/hour NAPHTHA FUEL
Catalytic desulfurizer	40	
Hydrogen production	58	
Steam generation(3)	338	
GROSS TOTAL	436	22,947
Supplied by process fuel gas(4)	-73	-3,847
NET TOTAL	363	19,100

---

(1) Naphtha heating value is 19,000 Btu/lb (HHV)
(2) SNG heating value is 1000 Btu/SCF (HHV)
(3) Includes superheating of steam
(4) By-product fuel gas produced and used in-plant

TABLE 4M -- THERMAL EFFICIENCY OF SNG PRODUCTION FROM NAPHTHA - METRIC UNITS

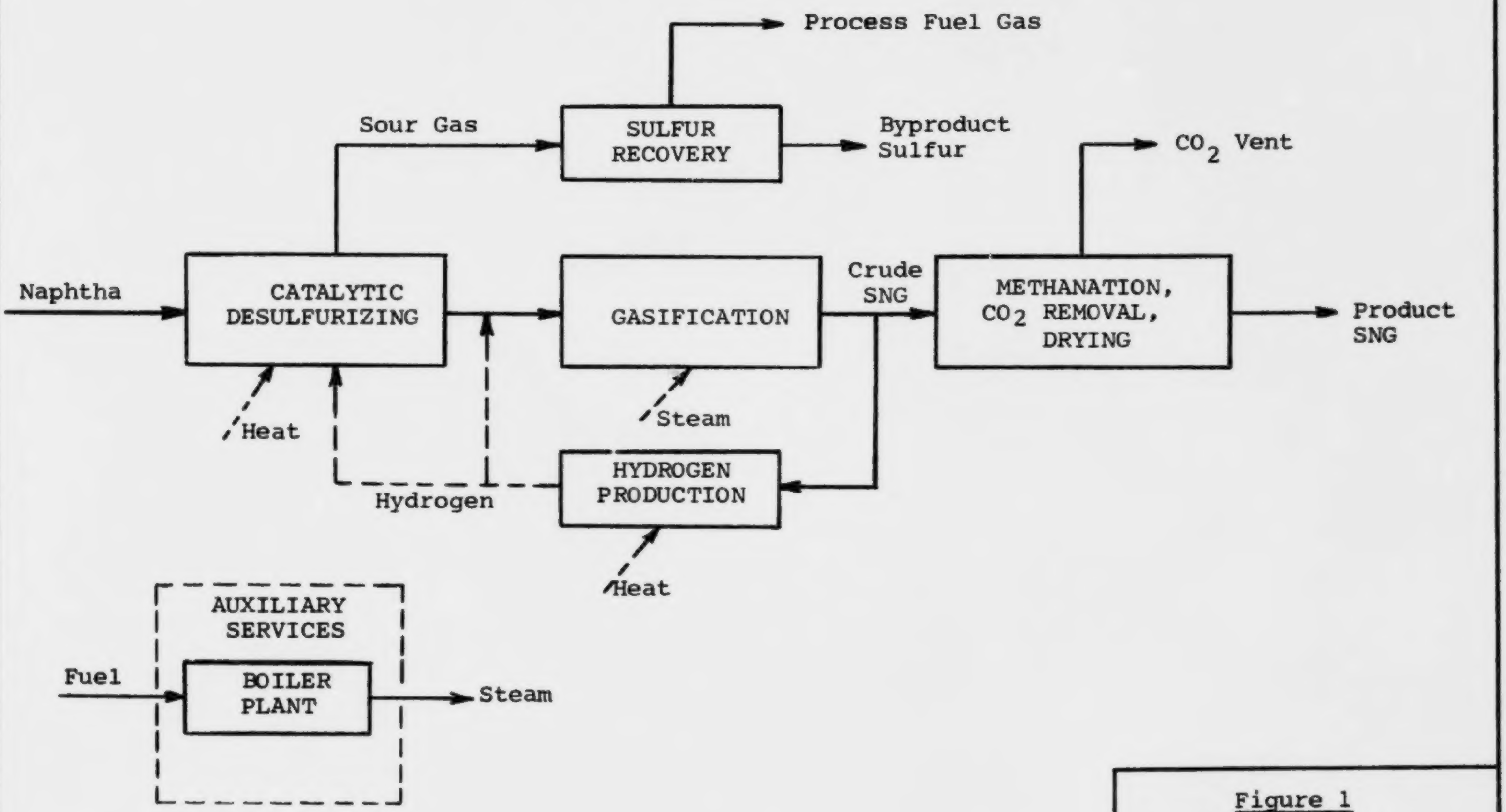
OVERALL THERMAL EFFICIENCY	FLOW RATE	Gcal/Day
Naphtha Feedstock, kg/hr	92,988	23,557(1)
, (m^3/day)	(3,013)	
Naphtha Fuel, kg/hr	8,664	2,195(1)
, (m^3/day)	(281)	
TOTAL INPUT		25,752
SNG Product, nm^3/day	2.679×10^6	25,200(2)

Thermal Efficiency = (25,200/25,752)(100)
= 97.8%

FUEL REQUIREMENTS	Gcal/hr HEAT RELEASE	kg/hr NAPHTHA FUEL
Catalytic desulfurizer	10.1	
Hydrogen production	14.6	
Steam generation(3)	85.2	
GROSS TOTAL	109.9	10,409
Supplied by process fuel gas(4)	-18.4	-1,745
NET TOTAL	91.5	8,664

---

(1) Naphtha heating value is 10,556 kcal/kg (HHV)
(2) SNG heating value is 9,406 kcal/nm^3 (HHV)
(3) Includes superheating of steam
(4) By-product fuel gas produced and used in-plant


Process Fuel Gas
Sour Gas
SULFUR RECOVERY
Byproduct Sulfur
CO_2 Vent
Naphtha
CATALYTIC DESULFURIZING
GASIFICATION
Crude SNG
METHANATION, CO_2 REMOVAL, DRYING
Product SNG
Heat
Steam
HYDROGEN PRODUCTION
Hydrogen
Heat
AUXILIARY SERVICES
Fuel
BOILER PLANT
Steam

Naphtha is used as feedstock, and as process heater and boiler fuel

Figure 1
Process Flow Diagram

SNG FROM NAPHTHA

FUEL REQUIREMENTS AND STACK GASES

Table 4 presents the fuel requirements as well as the overall thermal efficiency of the plant.

The total heat release required in the plant is 436 MM Btu/hr. We can estimate the unit stack gas rate, when burning naphtha, as about 11,000 SCF of WFG/MM Btu (see Table 3E). Therefore, the total stack gases issuing from the plant will be about 4,800,000 SCF/hr.

SULFUR BALANCE AND SO_2 EMISSIONS

20,700 barrels/day of feedstock plus fuel naphtha (with a sulfur content of 0.1 wt %) will contain 5,378 lbs/day of sulfur. Of this, 91.5% is recovered as by-product sulfur plus a small amount of waste zinc sulfide.

The remaining 8.5%, or 459 lbs/day, of sulfur leaves the plant as SO_2 in the stack gases from burning of the fuel naphtha in the process heaters and steam generator. Thus, the SO_2 emissions in the stack gases will be 918 lbs/day. This amounts to only 0.09 lbs SO_2/MM Btu of heat release, which easily meets the EPA limit of 0.8 lbs/MM Btu for firing liquid fuels.

The sulfur balance specifics are presented in Table 5.

WATER BALANCE

The water balance, shown in Table 6, is based on a design utilizing air-coolers for supplying all of the plant's cooling needs. The total supply water demand is 380 gpm for boiler feedwater. About 79% of this is converted to hydrogen for producing methane since steam is the source of hydrogen required for SNG production. Another 4.5% is vented with the CO_2 removed from the product SNG. The remaining 16.5% is discharged as effluent waters.

TABLE 5E -- SULFUR BALANCE FOR SNG PRODUCTION FROM NAPHTHA - ENGISH UNITS

	sulfur lbs/day	SO_2 emissions lbs/day
Feedstock Naphtha @ 0.1 wt % S	4920	-
Fuel Naphtha @ 0.1 wt % S	458	-
	5378	
By-product sulfur	4870	-
Desulfurizer heater stack gas	51	102(2)
Hydrogen plant heater stack gas	73	146(2)
Steam generation	335	670(2)
Waste zinc sulfide(1)	49	-
	5378	918

Overall sulfur emissions = (51 + 73 + 335)(100)/5378
= 8.5% of input sulfur

Overall fuel heat release (Table 4)
= 436 x 10^6 Btu/hr

Overall SO_2 emissions = (918/24)/436
= 0.09 lbs SO_2/MM Btu (HHV)

EPA SO_2 limit for firing liquid fuels
= 0.80 lbs SO_2/MM Btu (HHV)

---

(1) From use of zinc oxide to absorb final sulfur traces from feedstock naphtha after catalytic desulfurization

(2) One pound of sulfur is equivalent to two pounds of SO_2

TABLE 5M -- SULFUR BALANCE FOR SNG PRODUCTION FROM NAPHTHA - METRIC UNITS

	Sulfur kg/day	SO_2 emissions kg/day
Feedstock Naphtha @ 0.1 wt % S	2232	-
Fuel Naphtha @ 0.1 wt % S	208	-
	2440	
By-product sulfur	2209	-
Desulfurizer heater stack gas	23	46(2)
Hydrogen plant heater stack gas	33	66(2)
Steam generation	152	304(2)
Waste zinc sulfide	22	-
	2439	416

Overall sulfur emissions = (23 + 33 + 152)(100)/2439
= 8.5% of input sulfur

Overall fuel heat release (Table 4)
= 109.9 Gcal/hr

Overall SO_2 emissions = (416/24)/109.9
= 0.16 kg SO_2/Gcal (HHV)

EPA SO_2 limit for firing liquid fuels
= 1.44 kg SO_2/Gcal (HHV)

---

(1) From use of zinc oxide to absorb final traces of sulfur from feedstock naphtha after catalytic desulfurization

(2) One kilogram of sulfur is equivalent to two kilograms of SO_2

TABLE 6 -- WATER BALANCE FOR SNG PRODUCTION FROM NAPHTHA

	gpm	m^3/hr	%
INPUTS			
Boiler feedwater supply(1)	380	86.2	
OUTPUTS			
Conversion to methane	300	68.1	78.9
Vented with CO_2	17	3.9	4.5
Demineralizer blowdown	38	8.6	10.0
Boiler blowdown	12	2.7	3.2
Treated wastewater discharge(2)	12	2.7	3.2
On by-product sulfur	1	0.2	0.2
	380	86.2	100.0

Total eventually returned to atmosphere = 78.9 + 4.5
= 83.4%

---

(1) This plant was designed for total air-cooling, and hence no cooling water system makeup is needed

(2) This small amount could be vaporized in boiler fire-box and be rejected as water vapor in the boiler stack gases

As noted in Table 6, some of the effluent discharge could be vaporized in the boiler firebox if the situation required it.

SUMMARY OF MAJOR ENVIRONMENTAL FACTORS

Total naphtha (feed plus fuel)		20,700 barrels/day 224,100 lbs/hr
SNG product (@ 1000 Btu/SCF)		100MM SCFD
Overall Thermal Efficiency		97-98% (approx.)
Sulfur by-product		4900 lbs/day (approx.) 2.5 tons/day (approx.)
Total combustion heat release		436 MM Btu/hr
Total combustion stack gases		4.8 MM SCF/hr
Total SO_2 in stack gases		918 lbs/day
SO_2 per MM Btu's of heat release		0.09 lbs/MM Btu
Total raw water intake		380 gpm
Total effluent water discharge		62 gpm
CO_2 vent gases:	CO_2	175,800 lbs/hr (approx.)
	H_2O	8,500 lbs/hr

The major emissions and effluent discharges listed above are:

(a) Combustion stack gases

(b) CO_2 vent gases

(c) Effluent water discharges (blowdowns and wastewater)

OVERALL PROCESS MATERIAL BALANCE

	lbs/hr	
The overall process inputs are:		
Naphtha feedstock	205,000	
Water (converted to SNG hydrogen and CO_2 oxygen)	150,000	(300 gpm)
	355,000	
The overall process outputs are:		
Product SNG	176,000	(100×10^6 SCFD)
Vent CO_2	175,800	
By-product sulfur	200	
Process fuel gas	3,000	
	355,000	

This balance excludes the fuel naphtha which merely exits as stack gas, and the water supply which exits as effluent discharges or with the vent CO_2. There would be no purpose served in showing these non-process inputs and their disposition.

OTHER FACTORS

The combustion stack gases will contain nitrogen oxides (NO_x) as well as SO_2, but, with a naphtha or lighter fuel, the NO_x emissions should easily meet the Federal limit of 0.3 lbs/MM Btu for firing liquid fuels. There should be essentially no emission of particulates from a naphtha or lighter fuel.

Some nominal amount of electric power will most probably be purchased for lighting, instruments, and small motor drivers for air-cooler fans and pumps.

Feedstock and fuel storage tanks will be required to permit continuous operation in the event of a temporary transportation interruption.

There will be no unusual noise problems, and a realistic limit of 50-60 dBA at the plant property line should be attainable during normal operation.

Periodically, minor amounts of solid wastes will require disposal (zinc oxide spent to zinc sulfide, and spent catalysts). The zinc sulfide disposal may occur once a month or so, and the spent catalyst may occur once each 12-24 months.

A relatively large emergency flare stack (24-inch diameter by 200 ft. high) will be needed. When flaring at the maximum emergency rate, the flame may extend 250 ft. above the flare stack and may create noise levels of about 80-90 dBA at a distance of 1000 ft.

OTHER PLANT CONFIGURATIONS

While the above description serves to present a general picture of an SNG plant, other designs may vary quite widely from the one given here. For example:

- -- An LPG feedstock would very probably not require desulfurizing. This would decrease the plant heat and stack gas releases, and drastically change the sulfur balance. A sulfur recovery unit would probably not be included.
- -- A naphtha with a different sulfur content would also change the sulfur balance and SO_2 emissions.
- -- The use of water-cooling rather than air-cooling would change the plant water balance significantly.
- -- The use of the CRG or MRG process rather than Lurgi's process might also alter some of the environmental factors.

ADDITIONAL READING

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SECTION V

SNG FROM COAL

Coal is the United States' most abundant energy source, with enough proven reserves to meet our energy needs for hundreds of years. Faced with crude oil shortages and dwindling supplies of natural gas, we must utilize our vast coal energy reserves.

The energy in coal can be utilized in three ways:

-- Direct burning for residential and industrial heating
The transporting and distributing of coal to the end-use residential and industrial market would be very costly, as would the modification or replacement of home heating furnaces to handle coal. The environmental impact in terms of air pollution would be extremely high. Although widely practiced decades ago, the direct burning of coal to supply heat can no longer be considered an acceptable alternative.

-- The burning of coal to generate electrical power
This is a viable alternative, assuming power generation plants are provided with electrostatic precipitators and stack gas scrubbers to remove particulates (fly ash) and SO_2. However, the thermal efficiency of generating electricity from any fossil fuel is quite low, ranging from 35-42% for the best of designs. And even with 85% removal of SO_2 (via stack gas scrubbing), a large power plant will release to the air over 100 tons/day of SO_2. Nonetheless, our need for electrical power will mandate the use of coal to generate power until nuclear power or other alternatives are developed and accepted.

-- The conversion of coal into clean-burning SNG for residential and industrial heating
Coal gasification is about 70% thermally efficient, or twice as efficient as the direct burning of coal to produce electricity. Gasification permits the recovery and removal of most of the coal sulfur content at 99+% efficiencies. A coal

> gasification plant is therefore much more environmentally desirable than a coal-fired power plant (even when transmission and end use efficiencies are also considered).

The commercial gasification of coal has been practiced for at least 50 years. The low-pressure Winkler coal gasifiers and others date back to the 1920s. Most of the early gasification processes were concerned with producing low-Btu 'towns gas' for residential use, or synthesis gas for producing hydrogen, methanol and ammonia. More recently the Lurgi high-pressure gasification process (developed in Germany) has been used extensively on a large-scale commercial basis in 14 plants around the world. Altogether, Lurgi gasifiers have successfully handled 59 grades of coal -- including coke, anthracite, semi-anthracite and sub-bituminous coals.

There are many coal gasification research programs currently underway in the U.S. to develop more advanced gasifiers. All of these are still in laboratory or pilot plant development, and none will be available for commercial application within the next 2-5 years.

Many major gas utility companies in the U.S. have decided that the Lurgi gasification process is technically feasible and fully proven for commercial production of pipeline quality, high-Btu SNG. There are at least 9 major gasification projects underway in the United States, and only one of these is seriously considering a process other than Lurgi's process. Most of these will involve the strip-mining of relatively low-sulfur Western coals.

THE LURGI COAL GASIFICATION PROCESS

Figure 2 is a schematic flow diagram for a coal gasification plant utilizing Lurgi technology to produce high-Btu SNG from strip-mined coal in New Mexico.

> This description of a coal gasification plant is based on a specific coal supply, a specific site, and a specific design, and is not universally applicable. It is intended merely to illustrate the processes involved.

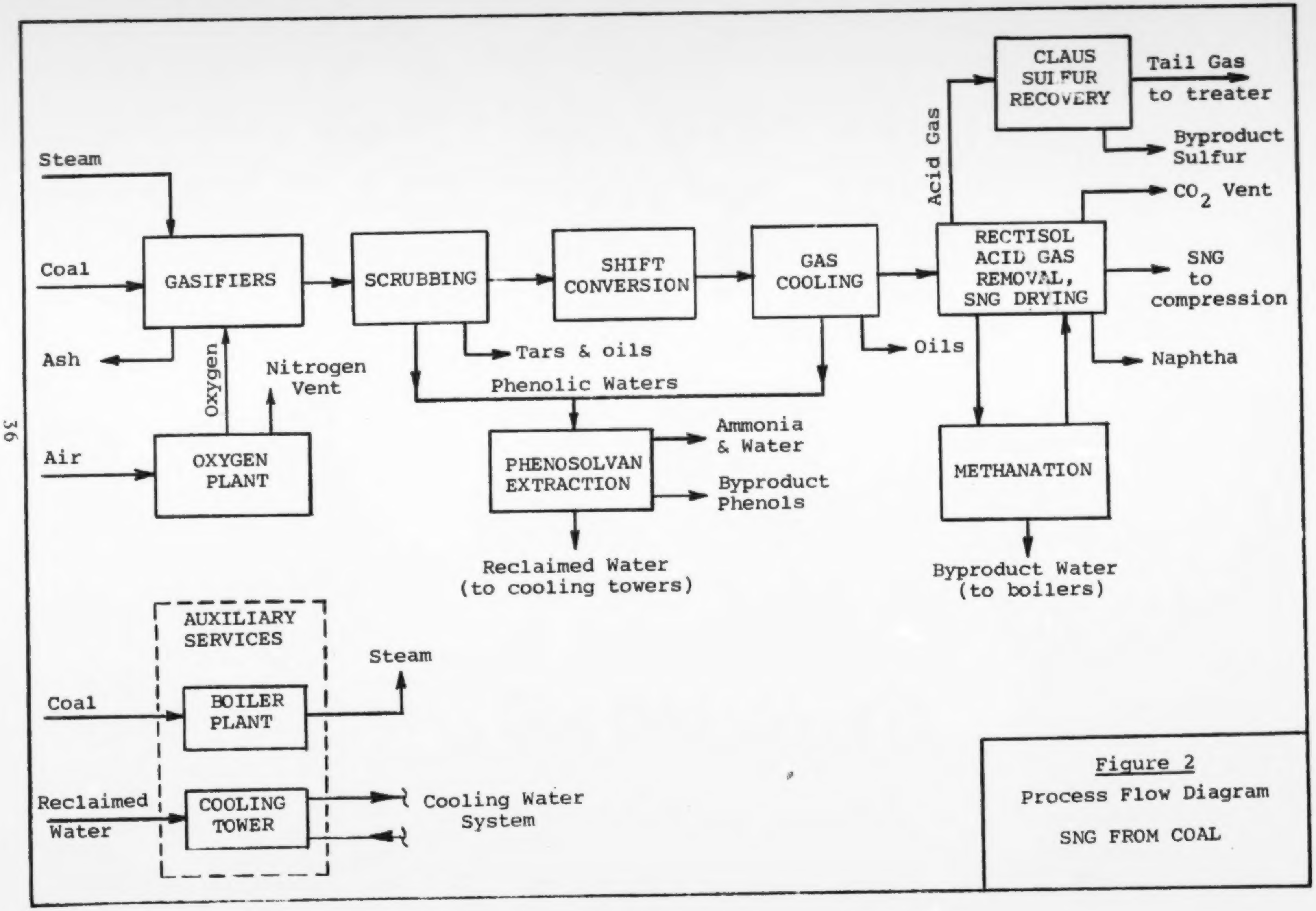
Steam
Coal
Ash
Air
GASIFIERS
Oxygen
OXYGEN PLANT
Nitrogen Vent
SCRUBBING
Tars & oils
SHIFT CONVERSION
GAS COOLING
Oils
Phenolic Waters
PHENOSOLVAN EXTRACTION
Ammonia & Water
Byproduct Phenols
Reclaimed Water (to cooling towers)
Acid Gas
CLAUS SULFUR RECOVERY
Tail Gas to treater
Byproduct Sulfur
RECTISOL ACID GAS REMOVAL, SNG DRYING
CO2 Vent
SNG to compression
Naphtha
METHANATION
Byproduct Water (to boilers)
AUXILIARY SERVICES
Coal
BOILER PLANT
Steam
Reclaimed Water
COOLING TOWER
Cooling Water System

Figure 2

Process Flow Diagram

SNG FROM COAL

The plant will produce 250 MM SCFD of SNG, as well as by-product fuels, from 25,600 tons/day of coal (about 9.4 MM tons of coal per year):

	Tons/day	10^6 SCFD	10^9 Btu/day
Gasification coal	21,860	-	-
Boiler plant coal	3,760	-	-
	25,620		442
Product SNG	-	250	250
By-product fuels	-	-	60
			310

Coal gasification involves a series of chemical reactions in which carbon from the coal is combined with hydrogen from steam to form methane, which constitutes 97 vol % of the product SNG. The heat required by the process is supplied by partial oxidation of the gasification coal with pure oxygen. Much of the heat is subsequently recovered by in-process generation of steam which is then reused as reaction steam and to supply equipment-driving energy (augmented by additional steam generated in auxiliary coal-fired boilers).

Very briefly, the reaction and conversion steps in Figure 2 include:

Pressure Gasification -- coal, steam and oxygen are reacted under controlled conditions of temperature and pressure to produce a crude gas containing methane, hydrogen, carbon monoxide, carbon dioxide, excess steam and various by-products and impurities. Only some 40% of the plant's end-product methane (SNG) is produced in the gasifiers. The remainder of the methane is produced in subsequent reaction steps.

Gas Scrubbing -- the crude gas is scrubbed and cooled with water, which removes tar and oil by-products and phenolic waters. The phenolic waters are subsequently processed for recovery of by-product phenols.

Shift Conversion -- excess carbon monoxide in the crude gas is 'shifted' (converted) to carbon monoxide to provide the 3-to-1 ratio of hydrogen-to-carbon monoxide needed for the subsequent synthesis of additional methane.

Gas Cooling -- the shifted gas is cooled again to remove additional hydrocarbon oil by-products and residual phenolic water.

Rectisol -- low temperature methanol is used to selectively absorb and remove H_2S and CO_2 from the cooled gas. Pre-cooling at the Rectisol unit entry also recovers by-product naphtha.

Methanation -- carbon monoxide and hydrogen are catalytically combined to produce methane and by-product water. About 60% of the end-product methane is produced in the methanation step.

Compression -- the dry, purified SNG is compressed and delivered to the pipeline with a heating value of 980-1000 Btu/SCF.

Phenosolvan -- a selective solvent (isopropyl ether) extracts by-product phenols from phenolic waters. The reclaimed water is stripped of H_2S and NH_3, and is further processed for complete reuse within the plant.

Claus Unit -- H_2S is catalytically converted to by-product sulfur and any residual gaseous sulfur compounds are incinerated to SO_2 and removed in a subsequent 'tail gas' treating unit.

Oxygen Plant -- pure oxygen is cryogenically extracted from atmospheric air.

Auxiliary Services -- these include coal-fired steam boilers with electrostatic precipitators to remove fly ash and stack gas scrubbers to remove SO_2. A closed loop, evaporative cooling water system as well as extensive air cooling is provided. Extensive wastewater treating and reuse is also provided.

OVERALL THERMAL EFFICIENCY

As tabulated earlier herein, the plant produces 310×10^9 Btu/day of SNG and by-product fuels from 442×10^9 Btu/day of gasification and boiler coal. This amounts to a 70% overall thermal efficiency. We could rationalize an even higher efficiency since some of the by-products (such as naphtha and phenols) will be sold for more than their mere heating value. We could thus argue that their contribution to overall energy recovery (thermal efficiency) should be higher since the income generated by their sales could be used to purchase more Btu's than they contain as heating value.

The only auxiliary fuel input required by the plant is that required to generate the auxiliary steam. Since the Lurgi pressure gasifiers must be fed coal of a certain size (ranging from about 3/16" minimum to 1½" maximum), the 'as-mined' coal must be crushed and screened to size. This produces a reject of smaller sized coal, referred to as 'coal fines'. The amount of fines produced fortuitously coincides (in this specific design) with the amount required for steam generation, so the design was based on burning coal fines to produce steam.

OVERALL PROCESS MATERIAL BALANCE

The overall material balance (not including the auxiliary steam boilers) for the gasification process producing 250 MM SCFD of SNG can be summarized as follows:

	Tons/day	wt %
INPUTS:		
Gasification coal	21,860	41.48
Steam and water	25,160	47.74
Oxygen	5,680	10.78
TOTAL	52,700	100.00
OUTPUTS:		
Product SNG	5,440	10.32
By-product phenols	105	0.20
By-product sulfur	175	0.33
Claus unit tail gas(1)	617	1.17
Ammonia plus water	800	1.52
By-product and reclaimed water	21,581	40.95
By-product hydrocarbon fuels	1,475	2.80
CO_2 vent gas	16,631	31.56
Gasification ash	5,876	11.15
TOTAL	52,700	100.00

(1) Does not include air used to incinerate the tail gas, and the material balance excludes the subsequent tail gas treating inputs and outputs.

The amount of coal in the above material balance for producing 250 MM SCFD of SNG is very specific to the particular coal being used.

For other coals, the amount may range from 18,000 to 36,000 tons per day. The steam and oxygen requirements may also vary over a wide range.

SULFUR BALANCE AND SO_2 EMISSIONS

The coal fed to the gasifiers has about 0.91 wt % sulfur and, therefore, the total sulfur contained in 21,860 tons/day of gasification coal is about 200 tons/day. Of that, about 16 tons/day will be retained in the naphtha and oil by-products and in the gasifier ash. The remaining 184 tons/day, in the form of gaseous H_2S, is processed in the Claus unit where about 175 tons/day is recovered as by-product sulfur. The last 9 tons/day of sulfur is incinerated to SO_2, of which 85% is removed as calcium sulfate solids in the tail gas treater. Finally, this leaves about 1.4 tons/day of sulfur released to the atmosphere in the form of SO_2, which is only 0.7% of the original 200 tons/day of sulfur in the gasification coal.

	Tons/day as Sulfur	Actual Form
INPUT:		
Sulfur in gasification coal	200	sulfur compounds
OUTPUTS:		
Sulfur in ash	10.0	sulfur compounds
Sulfur in tars, oils, naphtha	5.8	organic sulfur
Sulfur in CO_2 vent	0.5	carbonyl sulfide
By-product sulfur	174.5	sulfur
Calcium sulfate solids	7.8	$CaSO_4$ (gypsum)
SO_2 emissions to air from tail gas treater	1.4	SO_2
	200.0	

The coal fines, which are burned to generate steam, contain about 0.87 wt % sulfur after being washed to remove pyritic sulfur. The total sulfur contained in the 3,760 tons/day of coal required by the boilers is about 33 tons/day. Of that, about 5% remains in the boiler ash. The other 31+ tons/day becomes SO_2 in the boiler stack gases, where the stack gas scrubbers remove about 85% as calcium sulfate solids. Finally, this leaves about 4.7 tons/day of sulfur released to the atmosphere from the boiler plant in the form of SO_2.

The steam generated in the coal-fired boilers will be superheated in a separately fired unit using by-product oil containing about 0.9 tons/day of sulfur, all of which will be released as SO_2 to the atmosphere in the superheater stack gases.

	Tons/day as Sulfur	Actual Form
BOILER PLANT INPUTS:		
Sulfur in coal fines	32.7	sulfur compounds
Sulfur in by-product oil fuel	0.9	organic sulfur
	33.6	
OUTPUTS:		
Sulfur in ash	1.6	sulfur compounds
Calcium sulfate solids	26.4	$CaSO_4$ (gypsum)
SO_2 emissions to air:		
From boiler stack gas scrubbers	4.7	SO_2
From superheater	0.9	SO_2

In summary, the total emissions to the atmosphere from the gasification plant and the boiler plant will be 14 tons/day of SO_2 (the equivalent of 1.4 + 4.7 + 0.9 = 7.0 tons/day of sulfur) plus 0.5 tons/day of gaseous carbonyl sulfide contained in the CO_2 vent gas. Thus, only about 3% of the total 233.6 tons/day of sulfur entering the overall plant (gasification process plus boilers) is released to the atmosphere.

COMPARISON TO AN EQUIVALENT COAL-FIRED ELECTRIC POWER PLANT

As noted earlier in this section, the thermal efficiency of gasifying coal is twice as high as burning coal to generate electrical power (i.e. 70% versus about 35%). A power plant producing the same 310×10^9 Btu/day energy output as the gasification plant under discussion here would be generating 3780 MW and would have to use twice as much coal as the gasification plant, or about 51,000 tons/day. Assuming the power plant burns the same 0.91 wt % sulfur coal, its stack gases would contain about 920 tons/day of SO_2. Further assuming that the power plant included stack gas scrubbers to remove 85% of the SO_2, it would still release about 138 tons/day of SO_2 to the atmosphere, which is about 10 times greater than the 14 tons/day released by the equivalent coal gasification plant. The power plant is at a serious disadvantage in such a comparison because:

- -- The power plant uses twice as much coal
- -- The gasification plant removes the large bulk of process sulfur at 99+% efficiency and its boiler stack gas sulfur at 85% efficiency
- -- The power plant sulfur all issues from its boiler stacks and must all be recovered at the lower efficiency of 85%.

Direct burning is the most efficient way to produce electricity from coal. The above comparison is meant to illustrate the relative efficiency of producing energy via electricity generation versus via coal gasification -- for those residential and commercial uses where electricity and gas are in competition such as space heating, cooking, laundries, etc. In other words, coal can supply residential and commercial heating more efficiently via gasification than via electricity. As a corollary, the above comparison is not valid if the coal SNG is subsequently sold for use in generating electricity at 35% efficiency.

WATER BALANCE

Coal gasification requires large amounts of water as a source of hydrogen for producing methane SNG, for process cooling, and for generating steam energy. The gasification design under discussion will require about 5,100 gpm of raw water intake (8,200 acre-ft/year) to produce 250 MM SCFD of SNG. This amounts to 1.2 pounds of water intake[*] per pound of coal, including the boiler plant coal and including water used by the strip-mining operation (to be described later).

Figure 3 presents a schematic water flow diagram for the gasification plant, and Table 7 shows the overall disposition of the total raw water supplied to the plant.

Here again, as shown in Table 7, very little of the raw water intake is destroyed. In fact, about 80% of the water will be returned to the atmosphere, when we include the combustion water that will be returned wherever the product SNG is burned.

But the data in Figure 3 and Table 7 alone do not tell the full story of how much this design conserves water usage and maximizes the reuse of water. Some of the major features used to achieve those objectives are:

-- Steam turbines (driving process compressors) of about 250,000 horsepower will have air-cooled condensers as shown in Figure 3. If cooling water had been used instead, the additional evaporative loss alone would have totaled 4,000 gpm, almost doubling the plant water needs. As a point of interest, the air-cooled condensers provide about 2 billion Btu/hr of heat removal.

-- Much of the reaction steam supplied to the Lurgi gasifiers re-appears as phenolic wastewater (about 2500 gpm). By-product phenols are extracted from this water, dissolved gases are distilled out, and residual phenolics are biologically destroyed. The reclaimed water then supplies 100% of the cooling tower makeup needs. Finally, the cooling tower blowdown is again reused to quench the hot ashes from the Lurgi gasifiers.

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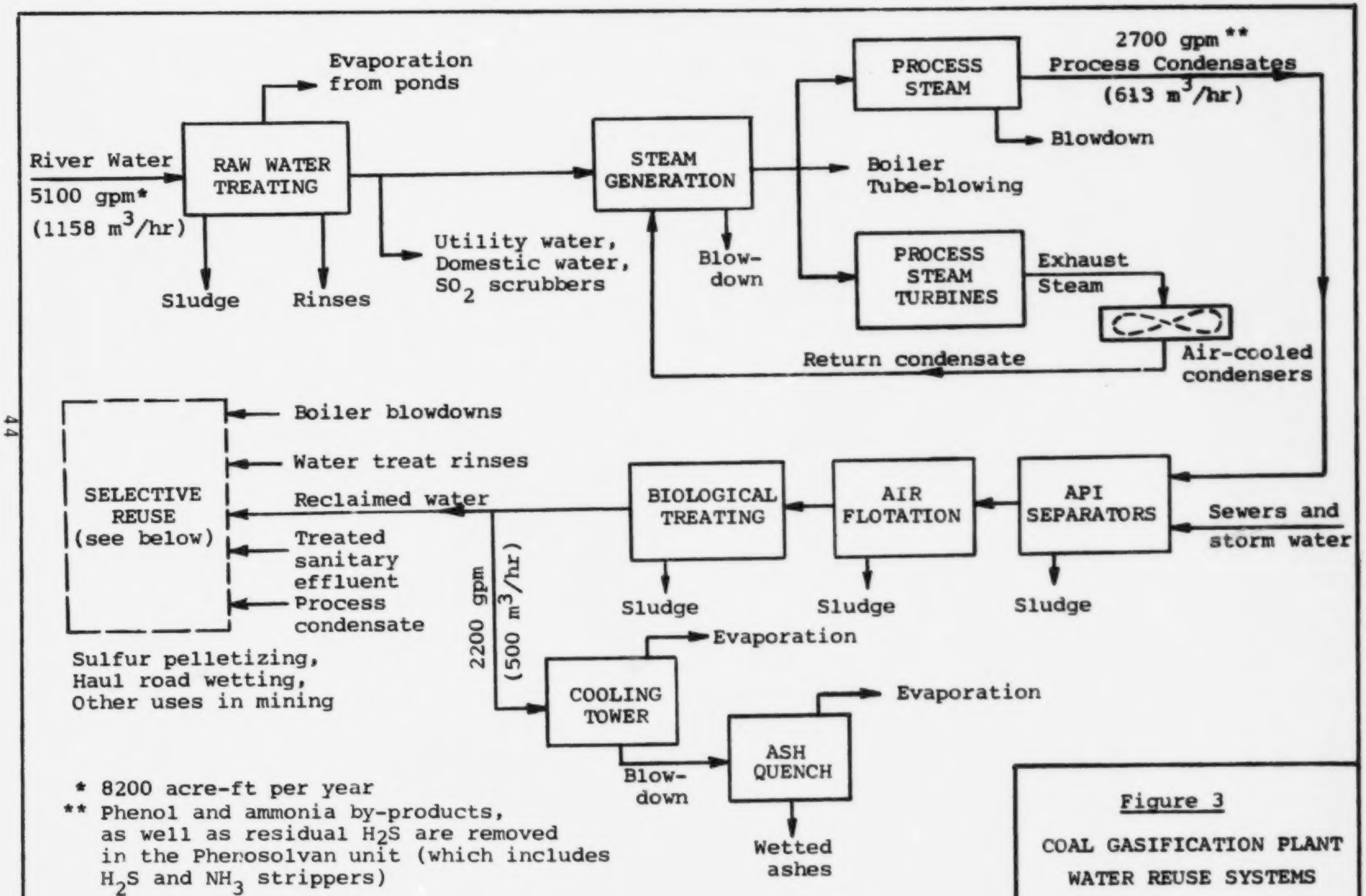
[*] This water requirement is very dependent on the particular coal involved and upon the specific plant design. It might range as high as 2 to 2.5 pounds per pound of coal.


River Water
5100 gpm*
(1158 m3/hr)
RAW WATER TREATING
Evaporation from ponds
Sludge
Rinses
Utility water, Domestic water, SO2 scrubbers
STEAM GENERATION
Blow-down
Boiler Tube-blowing
PROCESS STEAM
Blowdown
2700 gpm**
Process Condensates
(613 m3/hr)
PROCESS STEAM TURBINES
Exhaust Steam
Air-cooled condensers
Return condensate
SELECTIVE REUSE (see below)
Boiler blowdowns
Water treat rinses
Reclaimed water
Treated sanitary effluent
Process condensate
Sulfur pelletizing,
Haul road wetting,
Other uses in mining
BIOLOGICAL TREATING
AIR FLOTATION
API SEPARATORS
Sewers and storm water
Sludge
Sludge
Sludge
2200 gpm
(500 m3/hr)
COOLING TOWER
Evaporation
Blow-down
ASH QUENCH
Evaporation
Wetted ashes

* 8200 acre-ft per year

** Phenol and ammonia by-products, as well as residual H_2S are removed in the Phenosolvan unit (which includes H_2S and NH_3 strippers)

Figure 3

COAL GASIFICATION PLANT WATER REUSE SYSTEMS

TABLE 7 - WATER REQUIREMENTS AND DISPOSITION

	gpm	m^3/hr.	%
Process Consumption			
To supply hydrogen	1,120	254.2	
Produced as methanation by-product	-600	-136.2	
Net consumption	520	118.0	10.2
Return to Atmosphere			
Evaporation:			
From raw water ponds	420	95.3	
From cooling tower	1,760	399.5	
From quenching hot ash	150	34.1	
From pelletizing sulfur	250	56.8	
From wetting of mine roads	730	165.7	
	3,310	751.4	
Via stack gases(1):			
From steam blowing of boiler tubes	200	45.4	
From stack gas SO_2 scrubbers	40	9.1	
	240	54.5	
Total return to atmosphere	3,550	805.9	69.6
Disposal to Mine Reclamation			
In water treating sludges	100	22.7	
In wetted boiler ash	30	6.8	
In wetted gasifier ash	300	68.1	
Total disposal to mine	430	97.6	8.4
Others			
Retained in slurry pond	20	4.5	
Miscellaneous mine uses	580	131.7	
Total others	600	136.2	11.8
GRAND TOTAL	5,100	1,157.7	100.0

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(1) Does not include water derived from burning of boiler fuel

-- By-product water from the methanation synthesis is recovered for use as boiler feedwater.

-- Mechanical refrigeration is used in the Rectisol plant, rather than less costly absorption refrigeration, so that air-cooling could replace water cooling, thus avoiding evaporative water losses.

-- Finally, selected effluent waters will be used for pelletizing the by-product sulfur, and for dust abatement on the mining area roads.

There will be no discharge of effluent wastewater. Water not evaporated or converted to SNG is ultimately buried as wet ash and sludge in the strip-mine pits as they are filled and graded for land reclamation.

THE MINING OPERATION

The strip-mining of about 9.4 million tons of coal per year is a major operation. In fact, such a surface mine will be among the world's largest. It is beyond the scope of this report to go into any detail on the strip-mining, other than to emphasize its very large magnitude.

The task of filling the open pits, grading the land for reclamation, and providing the ultimate revegetation will be gigantic, and will be a significant factor in the environmental analysis of any coal gasification project.

STACK GASES

Table 8 is an itemized listing of the plant stacks, and the quantity and composition of their stack gas effluents.

OTHER ENVIRONMENTAL FACTORS

The key environmental factors, all discussed above, can be summarized

TABLE 8E -- STACK GASES FROM COAL GASIFICATION-ENGLISH UNITS

	Total Stack Gases				Tons/day of Emissions			Height,
	MM SCFD	tons/day	^oF	ft/sec	SO_2-	NO_x-	Partic.	feet
Claus tail gas	35	1,486	250	80-100	2.8	-	-	150-300
Boiler stacks	845	32,800	250	80-100	9.4	15.9	1.8	150-300
Superheater stack	138	5,370	600	80-100	1.7	1.5	0.026	150-300
Nitrogen vent	502	18,500	100	80-100	(almost pure nitrogen)			150-300
CO_2 vent	211	12,200	65	80-100	(0.5 tons/day of COS)			150-300
CO_2 vent	78	4,400	50	80-100	(99% CO_2 and 1% CH_4)			150-300
Coal lock vent	77	2,900	65	80-100	(99+% air, 10ppmv H_2S)			150-300
Ash lock vent	245	9,400	175	60-80	-	-	0.5	100-200
Coal conveyor vent	39	1,470	65	60-80	-	-	0.1	100-200
Emergency flare	(design data unavailable)				-	-	-	250-350

	Heat Release MM Btu/day	lbs/MM Btu Heat Release		
		SO_2-	NO_x-	Partic.
Claus tail gas	363,000$^{(1)}$	0.015	-	-
Boiler stacks	70,800$^{(2)}$	0.27	0.45	0.05
Superheater stacks	10,200$^{(3)}$	0.33	0.29	0.005

---

(1) Based on heating value of coal fed to Lurgi gasifiers
(2) Based on heating value of coal fines burned in boilers
(3) Based on heating value of by-product oils burned in superheater

TABLE 8M -- STACK GASES FROM COAL GASIFICATION - METRIC UNITS

	Total Stack Gases				Mg/day of Emissions*			Height,
	Mnm^3/day	Mg/day*	°C	m/sec	SO_2–	NO_x–	Partic.	metres
Claus tail gas	0.9	1,348	121	24-30	2.5	-	-	46-91
Boiler stacks	22.6	29,756	121	24-30	8.5	14.4	1.6	46-91
Superheater stack	3.7	4,872	316	24-30	1.5	1.4	0.024	46-91
Nitrogen vent	13.5	16,783	38	24-30	(almost pure nitrogen)			46-91
CO_2 vent	5.7	11,068	18	24-30	(0.45 Mg/day of COS)			46-91
CO_2 vent	2.1	3,992	10	24-30	(99% CO_2 and 1% CH_4)			46-91
Coal lock vent	2.1	2,631	18	24-30	(99+% air, 10ppmv H_2S)			46-91
Ash lock vent	6.6	8,528	79	18-24	-	-	0.45	30-61
Coal conveyor vent	1.0	1,334	13	18-24	-	-	0.09	30-61
Emergency flare	(design data unavailable)				-	-	-	76-107

	Heat Release	kg/Gcal Btu Heat Release		
	Gcal/day	SO_2–	NO_x–	Partic.
Claus tail gas	91,476(1)	0.027	-	-
Boiler stacks	17,842(2)	0.49	0.81	0.09
Superheater stacks	2,570(3)	0.59	0.52	0.009

---

(1) Based on heating value of coal fed to Lurgi gasifiers
(2) Based on heating value of coal fines burned in boilers
(3) Based on heating value of by-product oils burned in superheater

* Mg/day is equivalent to metric tons/day

very briefly as follows:

- -- Thermal efficiency and comparison to the alternative use of coal for power generation
- -- SO_2 emissions to the air
- -- Water consumption and disposition
- -- The mining operation and subsequent land reclamation

Other environmental factors are relatively minor by comparison, but are briefly discussed in this sub-section.

The plant will require about 30 MW of electric power for lighting, instruments, air-cooler fans, pumps and other uses.

All wetted ashes, water treatment sludges, and blowdowns are ultimately disposed of in the mining operation for land reclamation. The specific process proposed in this design for sulfur plant tail gas treating and boiler stack gas scrubbing is the Chiyoda Thorobred Process developed in Japan. It converts SO_2 to dry calcium sulfate ($CaSO_4$) as noted earlier. The $CaSO_4$ is of high enough quality to perhaps find a market in competition with natural $CaSO_4$ (gypsum). Other SO_2 removal processes may or may not produce a marketable end-product. In any event, if the selected process produces a 'throw-away' end-product sludge, it could be disposed of in the mining land reclamation along with the wetted ashes and water treatment sludges.

With 250,000 horsepower of steam turbines and a multitude of air-cooler fans, in-plant noise will be a distinct problem but not an insurmountable one. A limit of 60-70 dBA at the plant property line should be realistically attainable.

A good many storage tanks for chemicals, catalysts and liquid by-products will be required. Coal and sulfur by-product storage piles will also be needed.

There should be no problem with the periodic disposal of spent catalysts via burial in the strip-mine pits.

As can be noted from Table 8, electrostatic precipitators on the boiler stacks and other dust control measures will provide particulates emission levels (lbs/MM Btu of heat release) that will satisfy any anticipated regulations.

A very large emergency flare system will be required. When flaring at maximum emergency conditions, the flame will be quite high and very noisy. This condition, however, should occur only rarely.

Coal gasification plants will be located at the 'mine mouth', i.e. adjacent to the coal fields, and should therefore be in fairly remote sites. Such sites will probably require the concurrent construction of:

- -- Access roads and perhaps railroads
- -- Water supply pipelines
- -- Gas product pipelines

The plant and mine will require a total operating staff of perhaps 900-1000 people who, with their families, will have a permanent impact on local housing. The peak construction staff will number about 3500 personnel but their impact on local housing should extend over a 2-3 year period only. These operating and construction personnel will of course create a number of socio-economic impacts other than housing, which must be evaluated.

OTHER PLANT CONFIGURATIONS

Other design configurations may vary significantly from that described above. In particular, the choice of whether to burn coal fines to produce steam (as described herein), or whether to gasify the fines to produce low-Btu gas for steam generation or for compressor drive energy is one which is very difficult to assess. Whether to utilize very extensive air-cooling as in the specific design discussed herein, or whether to use less costly water-cooling, is another difficult choice to face. Finally, coals of other sulfur and ash contents as well as other heating values would result in different emission spectrums.

Although most of the coal gasification projects now underway in the U.S. are planning to use Lurgi process technology, this may not always be the case. Two other processes, dating back to the older low-Btu 'towns gas' era, have about the same amount of commercial experience as the Lurgi process. These are the Koppers-Totzek gasifiers with about 16 commercial installations, and the Winkler gasifiers also with about 16 commercial installations. Both of these processes were developed to operate at essentially atmospheric pressure, as contrasted with the Lurgi gasifier's operating pressure of 350-450 psi. The licensors of these two processes have not developed the up-to-date methanation technology required to produce pipeline quality, high-Btu SNG. Lurgi has developed the required methanation technology and is willing to guarantee their high-Btu SNG designs. Lurgi was also quicker to exploit the U.S. market and has gained a considerable advantage by being selected for most of the 'first-generation' SNG projects. This advantage may prove to be transient and, with certain types of coal, either Koppers-Totzek or Winkler may yet become a factor in the production of high-Btu SNG from coal.

In addition to the Lurgi, Koppers-Totzek and Winkler processes, there are a whole host of 'second-generation' process research programs underway in the U.S. to develop more advanced gasifiers. This is a partial listing of those programs:

Atgas (Applied Technology Corp.) -- demonstrated in small scale, short duration, batch tests

Bi-Gas (Bituminous Coal Research) -- 120 ton/day pilot plant under construction at Homer City, Pa.

CO_2 Acceptor (Consolidation Coal Co.) -- 40 ton/day pilot plant experiencing initial startup problems at Rapid City, S.D.

Hydrane (U.S. Bureau of Mines) -- 200 lbs/day bench scale unit in operation

Hygas (Institute of Gas Technology) -- 75 ton/day pilot plant is in operation at Chicago, Ill. Demonstration plant for 80 MM SCFD of SNG is in design

Molten Salt (M.W. Kellogg Co.) -- basic reactions established in laboratory

Synthane (U.S. Bureau of Mines) -- 70 ton/day pilot plant under construction at Bruceton, Pa.

Union Carbide Coal Gasification -- 25 ton/day pilot plant under construction at West Jefferson, Ohio

Westinghouse Coal Gasification -- 15 ton/day pilot plant under construction at Walez Mill, Pa.

Wellman-Galusha Gasification -- some commercial operation in the U.S. on a small scale to produce low-Btu gas.

These development programs probably will not result in a commercially viable process within the next few years. An optimistic estimate of the time required to complete these programs might be 2-5 years, and a pessimistic estimate might be 5-10 years.

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SECTION VI

SNG FROM CRUDE OIL

Basically, SNG is produced from crude oil by first producing naphtha and LPG, and then converting those products into SNG (as described in Section IV herein). However, it would be too difficult and too costly to so convert all the crude oil. The more reasonable approach is to process the crude oil in a refinery configuration that produces an SNG feedstock (naphtha and LPG) as well as a low-sulfur fuel oil. Most designs for producing SNG from crude oil are based on that approach. Plants producing SNG and low-sulfur fuel oil are called 'SNG refineries'.

As discussed in previous sections, all SNG plants require a source of hydrogen to combine with the feedstock's carbon -- whether from LPG, naphtha or coal -- and form methane SNG. Since the SNG refinery must first produce as much naphtha and LPG as reasonably possible, the refinery also requires a source of hydrogen because the crude oil has too low a hydrogen-to-carbon ratio. Finally, hydrogen is needed to convert the sulfur content of the crude oil into gaseous H_2S (acid gas) which can then be removed and converted into by-product sulfur. Therefore, all SNG refinery configurations include a hydrogen plant to:

-- Convert heavy hydrocarbons in the crude oil into lighter hydrocarbons (LPG and naphtha)

-- Convert LPG and naphtha into SNG

-- Convert sulfur into gaseous H_2S

There are dozens of configurations which could be used in an SNG refinery. The final selection depends on the specific case and its economics, the desired ratio of product SNG to product fuel oil, the market picture for alternative by-products (e.g. petroleum coke rather than fuel oil), the desired sulfur content of the product fuel oil, and the individual preferences of the plant owners and designers.

The composition of the crude oil (i.e. hydrogen-to-carbon ratio) also has a strong influence on the SNG refinery configuration.

About the only common criteria among the many possible configurations is that they all require a hydrogen plant and they all maximize the reasonable yield of naphtha and LNG for subsequent conversion into SNG.

As discussed in the previous Naphtha SNG section, there are at least three process technologies available for the final conversion of LPG and naphtha into SNG: (1) Lurgi's Gasynthan process, (2) the British Gas Council's CRG process, and (3) the Japanese MRG process.

A TYPICAL SNG REFINERY PROCESS DESIGN

Figure 4 is a schematic flow diagram of a typical SNG refinery design. Very few, if any, SNG refineries have yet to be constructed, and Figure 4 is therefore based on a preliminary study design. It has been simplified as much as possible, but an SNG refinery is a complex plant and it is difficult to simplify without becoming meaningless.

> This description of an SNG refinery is based on the use of a specific crude oil, at a specific site, and a specific production ratio of SNG to fuel oil, and is not universally applicable. It is intended merely to illustrate the processes involved.

In this particular design, the feedstock is a blend of crude oil and condensate. The products are pipeline quality SNG and low-sulfur fuel oil (0.3 wt % sulfur).

The configuration used in this design (Figure 4) utilizes an 'all-hydrogen' refinery*. That is, hydrogen is used to: (1) desulfurize

---

* A 'partial-hydrogen' refinery would still use naphtha hydrotreating, but the hydrocracker might be replaced by a non-hydrogen fluid-bed catalytic cracker. Also, the residuum catalytic desulfurizer might be replaced by coking of the residuum. Again, many alternative options are available.

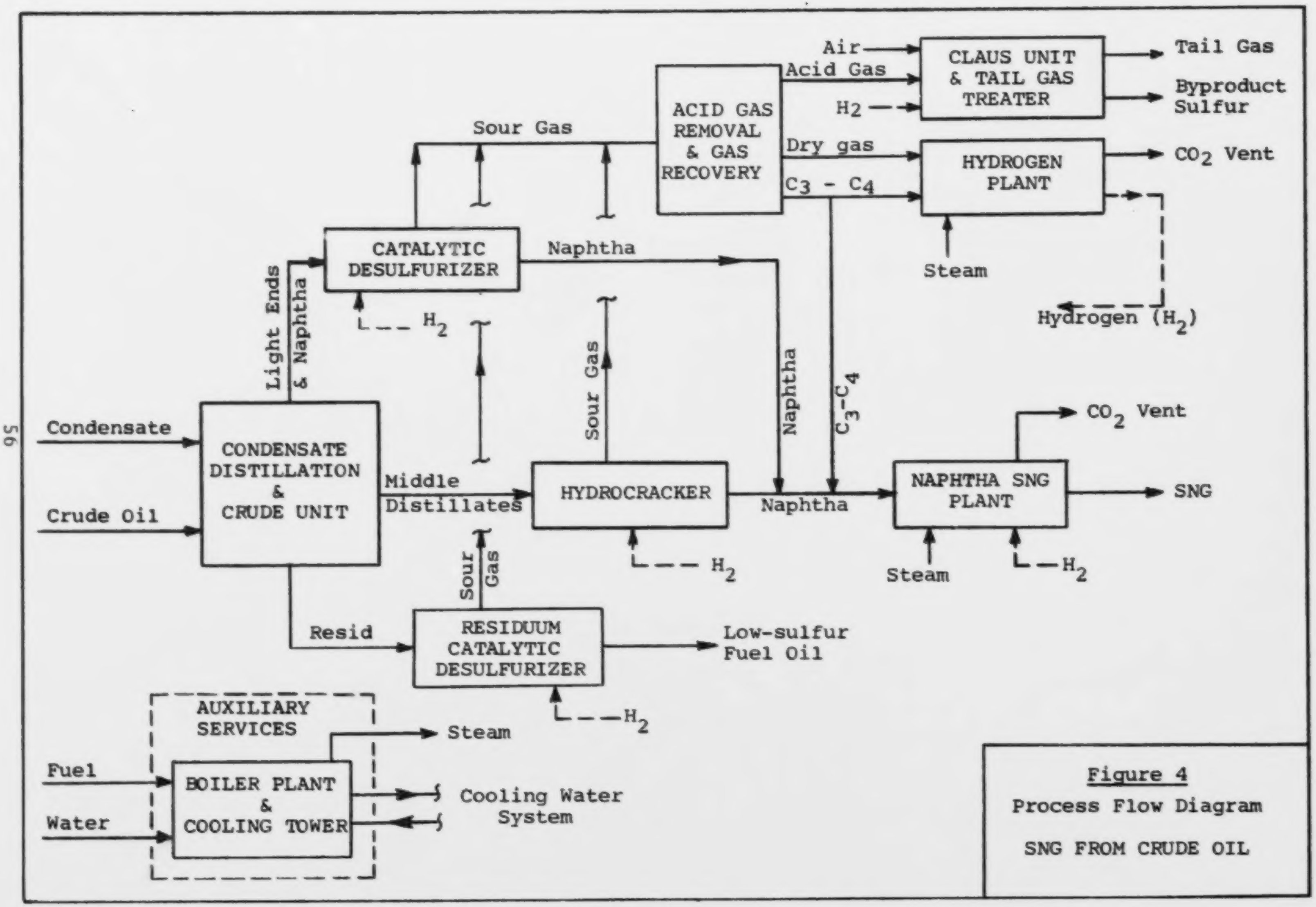
Condensate
Crude Oil
CONDENSATE DISTILLATION & CRUDE UNIT
Light Ends & Naphtha
CATALYTIC DESULFURIZER
H_2
Sour Gas
Naphtha
ACID GAS REMOVAL & GAS RECOVERY
Air
Acid Gas
H_2
CLAUS UNIT & TAIL GAS TREATER
Tail Gas
Byproduct Sulfur
Dry gas
$C_3 - C_4$
HYDROGEN PLANT
Steam
CO_2 Vent
Hydrogen (H_2)
Middle Distillates
HYDROCRACKER
H_2
Sour Gas
Naphtha
C_3-C_4
Naphtha
NAPHTHA SNG PLANT
Steam
H_2
CO_2 Vent
SNG
Resid
Sour Gas
RESIDUUM CATALYTIC DESULFURIZER
H_2
Low-sulfur Fuel Oil
AUXILIARY SERVICES
Fuel
Water
BOILER PLANT & COOLING TOWER
Steam
Cooling Water System

Figure 4
Process Flow Diagram
SNG FROM CRUDE OIL

or 'hydro-treat' the naphtha and the residuum fuel oil by converting sulfur to H_2S, and (2) crack or 'hydro-crack' the middle distillates into naphtha.

The overall thermal efficiency of the SNG refinery will be about 83%. It will produce some 1130 MM SCFD of SNG along with about 103,000 barrels per day of low-sulfur fuel oil from 365,000 barrels per day of feedstock:

	Barrels per day	Tons/day	10^6 SCFD	10^9 Btu/day
Crude oil	300,000	45,773	-	1,740
Condensate	65,000	8,300	-	337
	365,000			2,077
Product SNG	-	25,353	1,130	1,106
Product Fuel Oil	102,700	16,345	-	627
				1,733

A refinery feeding 365,000 barrels per day is not a small one -- it would rank among the largest oil refineries in the United States.

Very briefly, the processing steps in the SNG refinery (Figure 4) can be summarized in terms of the various unit processes:

Crude Unit -- In this unit, the crude oil and condensate are distilled to boil off a stream of light ends plus naphtha and a stream of middle distillate.

Residuum Desulfurizer -- The residuum from the crude unit distillation is catalytically desulfurized, using hydrogen to convert sulfur (in the residuum) into gaseous H_2S. This unit may also be called a hydrotreater.

Hydrocracker -- The middle distillate from the crude unit is catalytically cracked into smaller molecules of naphtha. Hydrogen is used to saturate the naphtha (i.e. provide the needed hydrogen-to-carbon ratio) and simultaneously convert sulfur in the middle distillate into gaseous H_2S.

Catalytic Naphtha Desulfurizer -- The virgin naphtha and light ends from the crude unit are catalytically desulfurized (again using hydrogen to convert sulfur into gaseous H_2S). This unit may also be called a hydrotreater.

Gas Recovery Unit -- The light ends and H_2S streams (sour gases) from the hydrocracker and the two desulfurizers are distilled in this unit. The resultant streams of dry gas (H_2, C_1, C_2) and C_3-C_4 then pass through an organic amine solution which absorbs and removes H_2S from those streams. The amine is subsequently boiled to release the H_2S (acid gas) which is sent to the Claus sulfur recovery unit. The dry gas is sent to the hydrogen plant as feedstock. A part of the C_3-C_4 LPG stream is also used as hydrogen plant feedstock, and the remainder is used as SNG feedstock.

Hydrogen Plant -- The dry gas and C_3-C_4 LPG feedstocks are converted into hydrogen, using steam to provide additional hydrogen.

Claus Unit -- Here the H_2S acid gas is burned with air and then catalytically converted to by-product sulfur. The residual 'tail gas' is then further processed, utilizing hydrogen once again, to recover additional by-product sulfur.

Naphtha SNG Plant -- Finally, the naphtha streams from the hydrocracker and from the naphtha desulfurizer, along with LPG from the gas recovery plant, are converted into SNG. Steam and hydrogen are used to combine with the hydrocarbons and to produce methane SNG and reject CO_2. (This is essentially the same unit as discussed previously in Section IV on Naphtha SNG. It includes steps for gasification, methanation of carbon monoxide, and the removal of CO_2 and water).

Auxiliary Services -- These include a boiler plant to provide process steam as well as steam for generating electrical power. The boilers will burn fuel produced in the refinery. A closed-loop, evaporative cooling water system is also provided, as are provisions for some air-cooling equipment.

OVERALL PROCESS MATERIAL BALANCE

The overall material balance for the SNG refinery can be summarized as follows:

	Tons/day	Wt %
INPUTS:		
Crude oil (less sulfur)	44,668	59.65
Condensate (less sulfur)	8,280	11.06
Sulfur	1,125	1.50
Air (process consumption only)	2,467	3.30
Steam	18,339	24.49
TOTAL	74,879	100.00
OUTPUTS:		
Product SNG	25,353	33.86
Product fuel oil	16,345	21.83
By-product sulfur	1,055	1.41
CO_2 vents	21,738	29.03
Tail gas	1,882	2.51
Fuel oil (consumed in-plant)	8,506	11.36
TOTAL	74,879	100.00

This balance does not include combustion air for heaters and boilers, or the resultant stack gases.

THERMAL EFFICIENCY COMPARISON FOR VARIOUS SNG PLANTS

As tabulated earlier in this section, the SNG refinery produces 1,733 x 10^9 Btu/day of product SNG and fuel oil from 2,077 x 10^9 Btu/day of crude oil and condensate, which is an overall thermal efficiency of 83.4%. It is interesting to compare the different thermal efficiencies of producing SNG and by-products from naphtha, from crude oil and from coal:

	Thermal Efficiency
SNG from Naphtha (Section IV)	97.8%
SNG from Crude Oil (Section VI)	83.4%
SNG from Coal (Section V)	70.0%

As would be expected, coal is the feedstock with the lowest hydrogen-to-carbon ratio and is the most difficult to gasify. Naphtha has the highest hydrogen-to-carbon ratio and is the easiest to gasify.

SULFUR BALANCE AND SO_2 EMISSIONS

The crude oil feedstock has a sulfur content of 2.46 wt %, and the condensate feedstock contains essentially no sulfur (0.0025 wt %). This amounts to a total sulfur input of 1125 tons/day.

The plant sulfur balance can be summarized as:

	Tons/day as Sulfur	wt %	Actual Form
INPUT:			
Crude oil and condensate	1125	100.00	organic S
OUTPUTS:			
Sulfur in SNG product	nil	nil	-
Sulfur in fuel oil product	49.0	4.36	organic S
By-product sulfur	1055.0	93.78	sulfur
Treated tail gas	0.5	0.04	SO_2
Stack gases (from in-plant fuel)	20.5	1.82	SO_2
	1125.0	100.00	

The total sulfur emissions to the air amount to 21 tons/day, or 1.86% of the input sulfur. The equivalent SO_2 emissions are 42 tons/day.

WATER BALANCE

The SNG refinery, with its auxiliary boilers and cooling tower, will require 12,620 gpm of raw water intake. The overall water balance is summarized below:

	gpm	m^3/hr	%
INTAKE:			
Raw water	12,620	2,865	100.0
DISPOSITION:			
Process consumption (hydrogen supply)	3,057	694	24.2
Cooling tower evaporation	6,080	1,380	48.2
Vents	698	158	5.5
Treated effluent discharge	2,785	632	22.1
	12,620	2,864	100.0

Once again, we note that about 78% of the raw water intake is ultimately returned to the atmosphere, when we include the water of combustion that will be released wherever the product SNG and fuel oil are burned. This is compared with 83% for the naphtha SNG plant and 80% for the coal SNG plant.

The treated effluent discharge of 2,785 gpm includes boiler and cooling tower blowdowns, as well as treated effluent waters. A conventional refinery waste treatment plant processes these waters before discharge. Treatment includes:

- -- In-plant sewer segregation and sour water stripping.
- -- Primary oil and suspended solids removal in an API separator and air flotation unit.
- -- Secondary treatment via biological oxidation.
- -- Final disinfection.

STACK GASES

Table 9 is a listing of the plant stacks, and their stack gas effluents.

TABLE 9E -- STACK GASES FROM SNG REFINERY - ENGLISH UNITS

	Stack Gases		Tons/day			lbs/MM Btu	
	MM SCFD	tons/day	SO_2	NO_x	MM Btu/day	SO_2	NO_x
H_2 Plant Heaters	898	34,332	0.2	8.2	54,720	0.01	0.30
Boilers	2,678	102,400	29.1	26.5	176,360	0.33	0.30
Process Heaters	1,200	45,936	12.0	10.8	72,580	0.33	0.30
Treated Tail Gas	-	1,882	1.0	-	-	-	-
CO_2 Vents	375	21,738	-	-	-	-	-
			42.3	45.5			

(The H_2 Plant heaters use a very low sulfur in-plant intermediate fuel. Hence, the lower emissions of SO_2 on a lb/MM Btu basis)

TABLE 9M -- STACK GASES FROM SNG REFINERY - METRIC UNIT

	Stack Gases		Mg/day*			kg/Gcal Btu	
	Mnm^3/day	Mg/day*	SO_2-	NO_x-	Gcal/day	SO_2-	NO_x-
H_2 Plant Heaters	24.1	31,146	0.18	7.44	13,789	0.013	0.540
Boilers	71.7	92,897	26.40	24.04	44,443	0.594	0.540
Process Heaters	32.1	41,673	10.89	9.80	18,290	0.594	0.540
Treated Tail Gas	-	1,707	0.91	-	-	-	-
CO_2 Vents	10.0	19,721	-	-	-	-	-
			38.38	41.28			

(The H_2 Plant heaters use a very low sulfur in-plant intermediate fuel. Hence, the lower emissions of SO_2 on a kg/Gcal Btu basis)

* Mg/day is equivalent to metric tons/day

OTHER ENVIRONMENTAL FACTORS

The key environmental factors, which already have been discussed above, are:

- -- Water consumption and disposition
- -- SO_2 emissions

Other environmental factors, all relatively minor, are briefly discussed below.

There are a number of solid waste disposal problems involving raw and effluent water treatment sludges, spent catalysts and spent chemicals. The sludges may amount to as much as 40 tons/day, and the design proposes to dispose of them in a 20-30 acre biological land cultivation area. Spent catalysts and chemicals require disposal only at infrequent intervals (12-24 months), and these will either be returned to their manufacturers for reclaiming or, in some cases, may possibly be used for land fill.

The usual noise problems associated with a major oil refinery will be encountered. However, good design practice should make a 55-65 dBA limit at the plant property line realistically achievable during normal operation.

About 135 MW of electrical power will be required by the plant, but it will be generated on-site in this design.

Storage tanks with a capacity of about 15 million barrels (630 million gallons) will be required for feedstock, product fuel oil, and in-plant intermediate products.

A very large emergency flare system will be required, just as in the coal gasification plant.

Since the fuel oil produced and burned in-plant has a very low ash content (about 0.01 wt % or less), there will be no problem with stack gas particulates.

Construction manpower will peak at about 5000 personnel. Depending on the plant location, this may cause local temporary housing problems and other socio-economic impacts.

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SECTION VII

LNG -- LIQUEFACTION AT SOURCE

Natural gas can be liquefied by refrigerating it to approximately -259°F at atmospheric pressure. Liquefaction of the gas 'shrinks' about 625 cubic feet of gas into 1 cubic foot of liquid. This makes it economically feasible to transport natural gas from remote overseas sources to the domestic U.S. market. The liquefied natural gas (LNG) is transported from those remote sources in very large, refrigerated ships called LNG tankers or LNG carriers. Each tanker can carry an amount of LNG which, when regasified at the market terminal, will become 3 billion SCF of natural gas.

TYPICAL PROCESS DESIGN FOR LNG LIQUEFACTION

First, it must be understood that there is no typical design for liquefying natural gas. Any given design depends on a number of factors:

- -- The pressure and composition of the raw natural gas ia a very important factor which will determine much of the design configuration.
- -- A decision must be made whether to remove LPG and condensate, if any, from the raw gas at the liquefaction site or whether to remove those natural gas liquids (NGL) at the market terminal. (The usual decision has been to remove the NGL at the liquefaction site although import regulations or other factors may make it more desirable to ship the NGL to the market terminal with the LNG).
- -- The type of refrigeration process must be chosen. There are a number of processes to be considered but generally they fall into two categories: the classical 'cascade' system and the 'mixed component' system.
- -- The location and ambient temperature conditions of the site are also factors in the design.
- -- Finally, individual preferences of the plant owners and the designers will affect the design.

In general, however, all LNG liquefaction plants have certain common characteristics. Since the raw gas must be refrigerated to very low temperatures, it must first be 'treated' to remove acid gases (H_2S and CO_2) and water which would otherwise freeze and plug the refrigeration system equipment. In any event, they must be removed for another reason: to meet pipeline and end-use specifications at the marketing site.

Next, the NGL will be removed and purified (if the gas contains any significant amounts), assuming the decision were made to remove them at the liquefaction site.

Finally, the gas will be cryogenically processed (refrigerated to very low temperatures) and liquefied at about -259°F. The LNG product is then stored in large tanks until loaded onto the LNG tankers. The boil-off vapors from the storage tanks, caused by atmospheric heat flowing into the cold tanks, are returned to the cryogenic plant for re-processing. The vapors displaced from the tanker, when loading out LNG, are returned to the storage tanks to replace the volume of liquid loaded out.

In summary then, an LNG liquefaction plant will consist of:

- -- Raw gas treating to remove H_2S, CO_2 and water.
- -- Removal and recovery of natural gas liquids (NGL), if any.
- -- Cryogenic liquefaction of the gas.
- -- Storage of the LNG product, and facilities for loading LNG aboard tankers or carriers.

Figure 5 presents a process flow diagram for a plant which liquefies an average of 200 MM SCFD of natural gas.

> This description of an LNG liquefaction plant is based on a specific feed gas composition and a specific site, and is not universally applicable. It is intended merely to illustrate the processes involved.

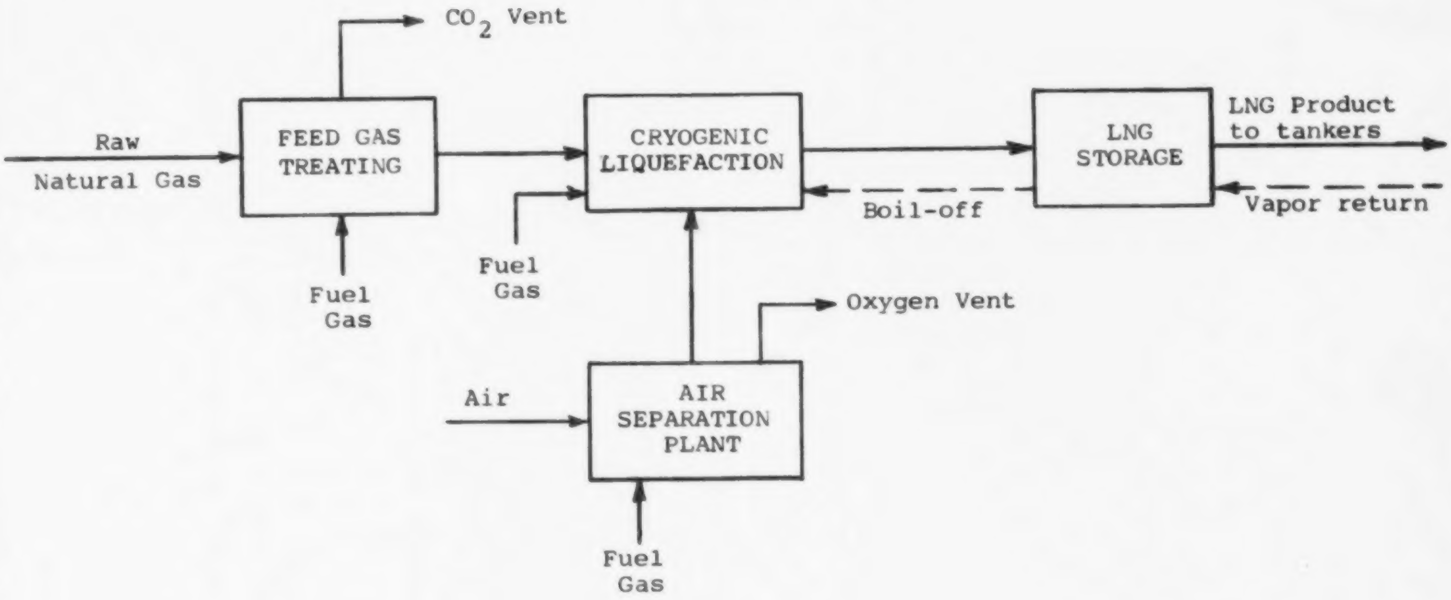
CO_2 Vent
Raw
Natural Gas
FEED GAS TREATING
Fuel Gas
CRYOGENIC LIQUEFACTION
Fuel Gas
LNG STORAGE
LNG Product to tankers
Vapor return
Boil-off
Oxygen Vent
Air
AIR SEPARATION PLANT
Fuel Gas

Figure 5

Process Flow Diagram

LNG LIQUEFACTION

This design was based on a very 'sweet' raw gas containing essentially no natural gas liquids, so the only raw gas treatment required is the removal of CO_2 and water. The compositions of the feedstock and the product are:

	Volume %	
	Raw Gas	LNG Product
Nitrogen	0.43	0.28
Methane	99.51	99.66
Ethane and Propane	0.06	0.06
	100.00	100.00
Water, ppmv	275	less than 1
CO_2, ppmv	1,000	less than 50
H_2S, ppmv	nil	nil

The plant was designed to utilize gas-turbines for driving the compressors in the refrigeration cycle. Since the chosen cycle was 'mixed component' refrigeration, and since one of the refrigerant components used in this case is nitrogen, an air separation plant is needed to obtain nitrogen from the atmosphere. Gas-turbines are also used for driving the compressors in the air separation plant. The total compression in the overall plant amounts to about 85-90,000 horsepower.

Air-cooling is used throughout the plant. All heating needs and all gas-turbines are supplied with treated feed gas as fuel, so no steam boiler is required. With no cooling tower and no boiler plant, the intake of raw water is very minimal, only 5,000 gallons per day (3.5 gpm).

THERMAL EFFICIENCY

The consumption of fuel for process and other heating and for the approximately 90,000 horsepower of gas turbines totals 18 MM SCFD of treated feed gas. Thus, the plant must feed 218 MM SCFD of gas to liquefy 200 MM SCFD. Therefore, the plant has an overall thermal efficiency of 91.7% based on feed and product heating values.

Some of the energy used in the plant, however, is provided by the pressure of the raw feed gas. If the gas enters the plant at 500-1000 psig and leaves as a liquid at atmospheric pressure, that expansion energy has been utilized within the process. Eventually, at the LNG market regasification site, most of that pressure will have to be re-supplied for the distribution pipeline. This point is mentioned to emphasize that the thermal efficiency of an LNG liquefaction plant is affected to some extent by the pressure energy available in the feed gas.

STACK GASES AND EMISSIONS

The only stack gas emissions from the plant are those resulting from the burning of 18 MM SCFD of fuel gas. By far, the largest amount of that fuel will be used in the gas turbines. The fuel gas will be treated feed gas, or almost pure methane with no sulfur content. The only emission of concern, therefore, is NO_x. A well-designed gas turbine burning methane should not produce more than 0.25-0.35 lbs. of NO_x per MM Btu heat release.

Although gas turbines use 200-300% excess air, we can use Table 3 to estimate that the total stack gas rate (corrected to 15% excess air) will be 11,830 SCF/MM Btu when burning methane. Thus, we can arrive at:

Total fuel burned	18 MM SCFD of C_1
Total heat release	18 x 10^9 Btu/day
Total wet stack gas rate	213 MM SCFD (8,149 tons/day)
Total NO_x	2.3-3.2 tons/day (0.25-0.35 lbs/MM Btu)

The 218 MM SCFD of feed gas contains 1000 ppmv of CO_2, which is reduced to 50 ppmv in the feed gas treater. This amounts to a vent of:

CO_2 vent	207,000 SCFD 24,040 lbs/day 12 tons/day

Finally, the air separation plant will vent about 1.5 tons/day of oxygen to the atmosphere.

WATER BALANCE

The total water intake to the plant will be only about 5,000 gallons/day because the total plant cooling in this specific case is provided by air-coolers. If a cooling water system had been necessary or economic, the water intake would have been much higher.

All of the water is used for plant wash-down, sanitary uses, potable drinking water and make-up to the plant's self-contained fire-fighting or firewater system. A small, prefabricated waste-treatment unit will process the effluent water from the plant and return 5,000 gallons/day to the local waterway.

OVERALL MATERIAL BALANCE

	tons/day	wt %
INPUT:		
Raw feed gas	4,612.0	99.8
Air (to air separation plant)	7.9	0.2
	4,619.9	100.0
OUTPUTS:		
LNG product	4,220.0	91.3
Fuel (burned in-plant)	380.0	8.3
CO_2 vent	12.0	0.3
O_2 vent	1.5	-
N_2 losses and purging	6.4	0.1
	4,619.9	100.0

OTHER ENVIRONMENTAL FACTORS

Only a very few other factors need be considered, and they are briefly discussed below.

The plant will require a very large emergency flare system, much the same as in all the other plants discussed in this report.

With 85,000 to 90,000 horsepower of gas turbines, the plant will have a distinct noise problem. Nonetheless, a plant property line limit of 60-70 dBA should be attainable.

Tankage on the order of 4,000-5,000 barrels will be required to store the liquid refrigerants.

About 1,000,000 barrels of cold LNG storage will also be required. The safety aspects of that storage deserve serious study in each case.

Dredging of the harbor to accommodate LNG tankers (if that is required) will create some environmental concerns that should be carefully considered.

OTHER PLANT CONFIGURATIONS

A design using water-cooling and steam turbines would be quite different, in terms of water and fuel balance, than the one described herein. If the feed gas contained H_2S, that would also alter the emission factors given.

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SECTION VIII

LNG -- REGASIFICATION AT MARKET

Liquefied natural gas (LNG) is transported to market regasification plants in very large carriers. Each tanker can deliver as much as 855,000 barrels of LNG, which will become about 3 billion SCF of natural gas when regasified.

Basically, an LNG regasification plant simply heats and boils the LNG to reconvert it to natural gas for pipeline distribution. The plant will contain three functional facilities:

- -- A tanker docking and unloading facility
- -- Large LNG storage tanks
- -- A regasification facility wherein the LNG is vaporized (i.e. heated and boiled)

In general, LNG will be withdrawn from the storage tanks as needed to supply the plant's base-load output of gas plus any peak-load requirements during cold weather. The LNG is raised to pipeline pressure (500-1000 psig) by cryogenic pumps, and is then vaporized by exchanging heat with water or by fuel-fired heating equipment. Since the LNG receiving terminal is usually in a coastal harbor, seawater is readily available for vaporizing the LNG.

TYPICAL DESIGN FOR LNG REGASIFICATION

Figure 6 presents a schematic flow diagram for a typical LNG regasification plant.

> This description of an LNG regasification plant is based on a specific design for a specific site, and is not universally applicable. It is intended merely to illustrate the processes involved.

This plant was designed to vaporize a base-load of about 1,000 MM SCFD of gas. It also includes facilities to vaporize an additional 450 MM SCFD of gas for 20 days per year (the peak-load days).

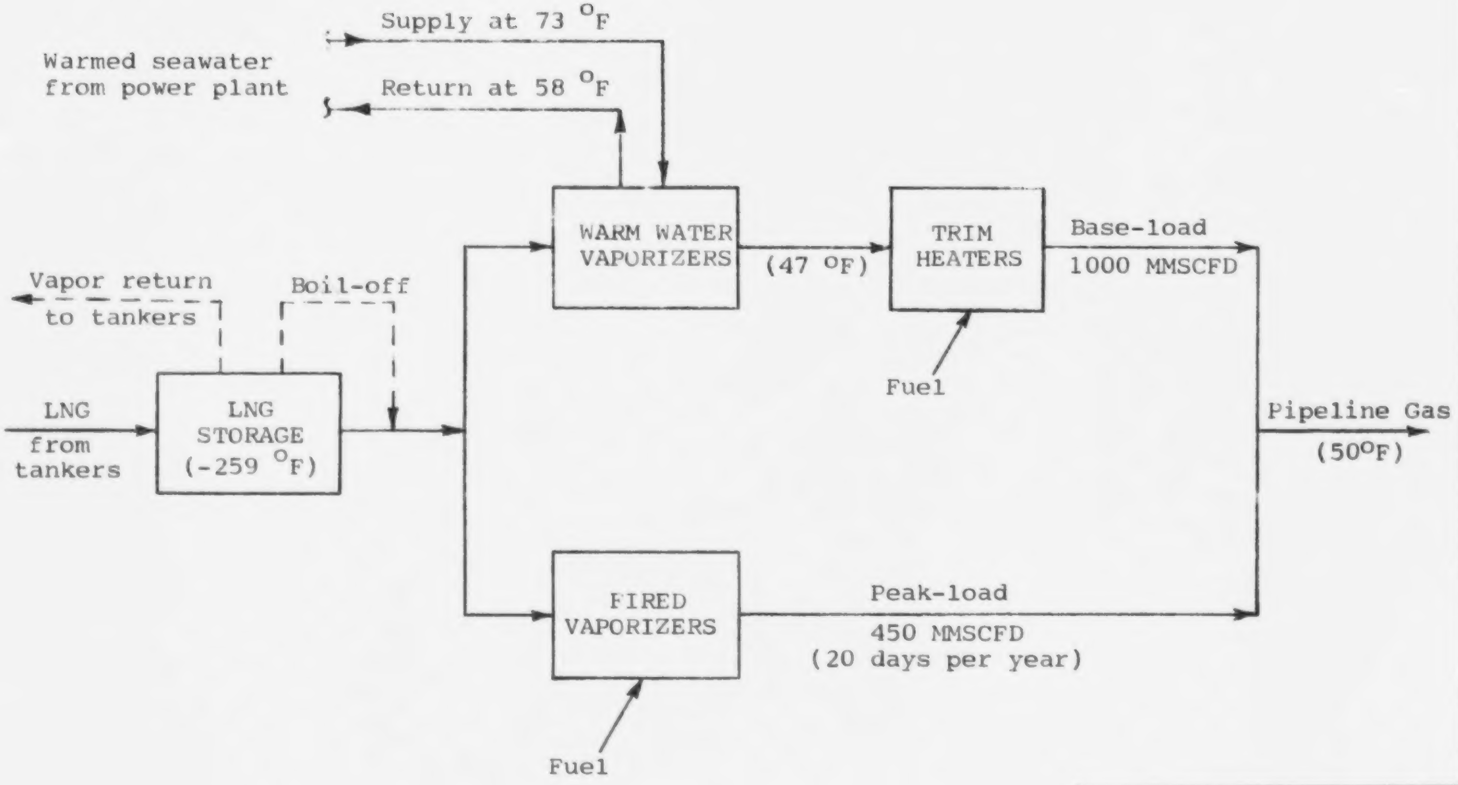
Supply at 73 °F
Warmed seawater from power plant
Return at 58 °F
WARM WATER VAPORIZERS
(47 °F)
TRIM HEATERS
Base-load
1000 MMSCFD
Fuel
Vapor return to tankers
Boil-off
LNG from tankers
LNG STORAGE (-259 °F)
Pipeline Gas
(50°F)
FIRED VAPORIZERS
Peak-load
450 MMSCFD
(20 days per year)
Fuel

Figure 6
Process Flow Diagram
LNG REGASIFICATION

Base-load vaporization in this design is accomplished by using warmed seawater (73°F) obtained from an adjacent large power plant*. The LNG and the seawater exchange heat in tubular exchangers wherein the LNG is heated and vaporized and the seawater is cooled. About 16 billion Btu's of heat per day are exchanged when vaporizing 1,000 MM SCFD of gas. The vaporized LNG leaves the exchangers at 47°F and is then 'trim-heated' to 50°F (required for pipeline transmission) by small gas-fired heaters, which supply about another 89 million Btu's per day.

The peak-load vaporization of an additional 450 MM SCFD of gas is accomplished in gas-fired vaporizers which will supply about 7 billion Btu's of heat per day.

The vaporizing operations can be summarized as follows:

	Base-Load	Additional Peak-Load (20 days/year)
Water:		
Inlet temperature, °F	73	-
Outlet temperature, °F	58	-
Flow rate, gpm	89,500	-
Heat exchange, 10^6 Btu/day	16,200	-
Vaporized gas:		
Flow to pipeline, MM SCFD	1000	450
Temperatures:		
From water exchangers	47	-
Into pipeline	50	50
Fired heat input, 10^6 Btu/day	88.8	7340
Fired heat fuel, 10^6 Btu/day	110.4	7640

---

* The power plant takes in seawater at 50°F, uses it to condense turbine exhaust steam and discharges the water at 73°F. The LNG regasification plant receives the water at 73°F and cools it to 58°F by heat exchange with LNG. Thus, water returns to the sea about 8°F warmer than originally obtained.

THERMAL EFFICIENCY

During the long voyage in the LNG tankers, about 2.5% of the LNG boils off (due to the absorption of atmospheric heat) and is used as tanker fuel. This fuel consumption will be included in the following accounting of thermal efficiency for the regasification plant.

The regasification plant uses about 25,000 KW of electrical power, mostly for the cryogenic pumps (which supply the required pipeline gas transmission pressure) and for the seawater pumps. This amounts to $2{,}000 \times 10^6$ Btu/day of energy. Since that energy is generated in a power plant typically operating at about 33% efficiency, the regasification plant is in fact using about $6{,}000 \times 10^6$ Btu/day of equivalent fuel when it consumes 25,000 KW of power. This fuel consumption will also be included in the following efficiency accounting. Finally, the 20 days of additional peak-load vaporizing will be pro-rated over the year to arrive at a total daily average basis for thermal efficiency.

		Delivered to Pipeline		
	On Tankers	Base Load	Pro-rated Peak-Load	Total
Gas, MM SCFD	1051	1000	25	1025
				10^9 Btu/day
HEAT INPUTS:				
Gas loaded on tankers (as LNG)				1051.0(1)
Electrical power (equivalent fuel)				6.0
Fuel for fired vaporizers				0.5(2)
				1057.5
HEAT OUTPUTS:				
Gas to pipeline				1025.0(1)
Fuel for tankers (boil-off)				26.0
Consumed (to pressurize & vaporize LNG)				6.5
				1057.5

Thermal Efficiency (based on product gas output)
= (1025/1057.5) 100 = 96.9%

(1) Assuming a gas heating value of 1000 Btu/SCF as typical
(2) The 7.64×10^9 Btu/day of peak load vaporizing fuel amounts to 0.4×10^9 Btu/day when pro-rated to 365 days. The trim heaters for base-load use another 0.1×10^9 Btu/day.

If we refer to Section VII on LNG liquefaction and consider that plant's thermal efficiency of 91.7% as being typical, then we have an overall LNG project thermal efficiency amounting to 91.7% of 96.9%, or 88.9%.

Where has the 11.1% of the heat been 'lost'? In a number of places:

- -- Supplying energy to refrigerate and liquefy the LNG at the source liquefaction plant
- -- Supplying fuel for the LNG tankers
- -- Supplying energy to pump seawater through the LNG vaporizers
- -- Supplying energy to pump the LNG up to pipeline distribution pressure
- -- Supplying fuel for gas-fired LNG vaporizers

If we consider the overall LNG project (liquefaction, tanker transport and regasification) as a gas transmission system over a 5,000-6,000 mile distance, the overall energy loss of 11.1% can be rationalized as about 2% per 1000 miles, which is quite good.

STACK GASES AND EMISSIONS

The only source of air emissions in the regasification plant design (Figure 6) is the stack gas from the fired vaporizers. Since the fuel supply is vaporized LNG, the sulfur content of the stack gases is practically nil, but they will contain nitrogen oxides as shown below:

	Base-Load Trim Heaters	Peak-Load Vaporizers
Stack gases, tons/day	53	2990
, MM SCFD	1.4	78
NO_x, lbs/day	20	917
, lbs/MM Btu	0.18	0.12

These are relatively insignificant emissions, especially when we consider that the peak-load vaporizers are only used 20 days per year.

WATER BALANCE

The only water balance factor in the plant is the use of 89,500 gpm on a once-through basis for vaporizing the base-load LNG. The water enters the gasification plant at 73°F and leaves at 58°F, which is only 8°F warmer than when originally taken from the sea by the power plant.

The LNG plant will also require about 3,000 gallons/day (2.1 gpm) of city water for sanitation and potable water.

OTHER ENVIRONMENTAL FACTORS

In general, an LNG regasification plant is a very clean one, without any major air emissions or wastewater discharges. The once-through cooling of seawater will normally create a "cold thermal impact" problem when the seawater is returned, although in the specific case discussed herein, that problem was solved by re-using warm water from an adjacent power plant.

The primary concern with an LNG receiving terminal will be one of safety. Incoming tankers will be fully loaded with LNG which will be off-loaded and stored on-site in very large tanks -- perhaps 2 to 3 tanks of 500,000 to 750,000 barrels each. This will require careful and detailed evaluation of each specific site.

Noise from the cryogenic and seawater pumps should be a relatively minor problem.

A large emergency flare will be required, just as in all the previous plant discussions.

Dredging of the harbor to accommodate the LNG tankers (if that is necessary) will create some environmental concerns that require study.

THE DIFFERENCE BETWEEN BASE-LOAD AND PEAK-SHAVING LNG PLANTS

A gas distribution company must supply two types of demand. One is a year-round demand for an average or 'base-load' amount of gas. The other is an additional 'peak-load' demand, during the coldest winter season, for an incremental amount of gas over and above the base-load.

Many gas companies are having difficulty in obtaining enough domestic U.S. gas to supply their average base-load. Such companies are importing LNG from overseas, and regasifying the LNG at coastal receiving terminals. These are usually very large installations, including LNG tanker docks and unloading systems, and are called 'base-load LNG plants'. It is these base-load plants which have been described in Sections VII and VIII herein.

Other gas companies may have somewhat more than enough domestic gas to supply their base-load demands, but not enough to supply their incremental peak-load demand in the winter. Those companies have two options:

1 -- During the summer, they can divert excess gas supplies into underground caverns for storage until the winter. Then during the peak season, they can withdraw gas from the caverns.

2 -- If suitable underground areas are not available, excess summer gas can be withdrawn from their supply pipeline and liquefied by refrigeration. The liquified gas (LNG) can be stored in large tanks, and withdrawn for regasification during the winter peak season. Such an installation is called a 'peak-shaving LNG plant'.

The basic difference between the two types of LNG plants (base-load and peak-shaving) is that the peak-shaver is usually a fairly small plant other than for a large LNG storage capability. Peak-shaving plants are smaller than imported LNG base-load plants because:

-- The incremental peak-gas demand may be only 10-40% of the base-load demand.

-- The peak season may only last 1-2 months, whereas the low-demand summer season may last 5-7 months. Thus, there may be 7 summer months in which to produce and store LNG for use during a 1 month peak season. This is a production-to-usage time ratio of 7:1. If the peak demand rate is 35% of the base-load rate, LNG can be produced and stored in the summer at a rate that is only 5% of the base-load rate.

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SECTION IX

METHANOL FUEL

As discussed in previous sections, natural gas can be liquefied for transport in large refrigerated LNG tankers to regasification terminals near the end-use market. As a technologically viable alternate, the natural gas could be converted to methanol (an alcohol) which can be transported to market and burned directly as a liquid fuel without any regasification step at all. Since methanol has a boiling point of 148°F, it is a liquid at ordinary temperatures and requires no refrigeration. Obviously, the cost of storage, handling and transportation facilities for methanol would be much less expensive than those for LNG. Because of this, many companies are evaluating the economics of transporting overseas natural gas as methanol rather than LNG. In fact, some companies have announced plans to proceed with such projects.

Some of the physical properties of methanol and LNG are compared below:

	Methanol(1)	LNG(2)
Chemical formula	CH_3OH	CH_4
Liquid density, lbs/ft^3	49.6(3)	26.4(4)
HHV, Btu/lb	9,750	23,900
HHV, Btu/CF of liquid	484,000	632,000
Boiling point at atmospheric pressure	148°F	-259°F

(1) Based on pure methanol
(2) Based on pure methane
(3) At 60°F
(4) At -259°F

TANKER AND CAPITAL REQUIREMENTS COMPARED TO LNG

Most published studies agree that the economic advantages of methanol storage and transportation make it competitive with LNG when the round-trip transportation distance is about 8,000-10,000 nautical miles. At even longer distances, a methanol project seems to be distinctly more economic than a comparable LNG project.

Table 10 presents a detailed comparison of the tanker requirements for equivalent methanol and LNG projects. Briefly, Table 10 can be summarized as follows:

	Methanol	LNG
Delivered product	25,000 tons/day	500 MM SCFD
Delivered energy, 10^{12} Btu/year	178	184
One-way trip, nautical miles	6,250	6,250
Round-trip, nautical miles	12,500	12,500
Tanker capacity	160,000 tons	125,000 cu.m.
Tanker speed	16 knots	20 knots
Round-trips/year/tanker	9.7	11.8
Tankers required	6	6

Thus, 6 tankers would be required for either a 25,000 ton/day methanol project or for a 500 MM SCFD LNG project, both large projects delivering essentially equivalent amounts of energy (178 to 184 x 10^{12} Btu/year).

160,000-ton methanol tankers would cost about $33 million each, and 125,000 cubic meter LNG tankers would cost about $80 million each. Therefore, the total tanker cost of the methanol project would be about $280 million less than the equivalent LNG project, based on mid-1973 costs and on a 12,500 nautical mile round-trip distance.

Comparing the overall capital investments for the equivalent plants, it can be seen that the savings in tanker costs more than offset the higher cost of the overseas methanol conversion plant relative to the LNG plant:

	CAPITAL INVESTMENT	
	Methanol	LNG
Delivered product	25,000 tons/day	500 MM SCFD
Capital investment:		
Overseas conversion plant	375 MM $	260 MM $
6 tankers	200 MM $	480 MM $
Market terminal	20 MM $	80 MM $
	595 MM $	820 MM $

These capital investment estimates are based on mid-1973 costs and are perhaps accurate within ± 20%. The LNG project includes regasification, and the methanol project does not.

TABLE 10E -- TANKER REQUIREMENTS (METHANOL VS LNG) - ENGLISH UNITS

	METHANOL	LNG
HHV, Btu/lb	9,750	23,900(1)
HHV, Btu/CF (liquid)	484,000	632,000(2)
Typical tankers:		
Delivered cargo, cu.meters	-	125,000
, short tons	160,000	-
, 10^{12} Btu	3.12	2.79
Travel speed, knots	16	20
12,500 nautical miles, round-trip:		
Round-trip travel, days	32.6	26.0
Loading/unloading/bad weather, days	5.0	5.0
Total trip, days	37.6	31.0
Round-trips per year	9.7	11.8
Energy delivered per tanker per year, 10^{12} Btu's	30.3	32.9
Project sizes & tanker requirements:		
Delivered product, tons/day	25,000	-
, MM SCFD	-	500
, 10^{12} Btu/year	178(3)	184(3)
Tankers required	6	6

---

(1) Equivalent to 1,010 Btu/SCF (gas)

(2) 1 cu. ft. of LNG liquid = about 625 cu.ft. of natural gas

(3) Practically equivalent project sizes since regasification of LNG consumes some energy for vaporizing and for pumping (see LNG Regasification section). Methanol sold as liquid fuel, with no regasification.

TABLE 10M -- TANKER REQUIREMENTS (METHANOL VS LNG) - METRIC UNITS

	METHANOL	LNG
HHV, kcal/kg	5,417	13,279 (1)
HHV, kcal/m^3 (liquid)	4,307,000	5,625,000 (2)
Typical tankers:		
Delivered cargo, cubic metres	-	125,000
, metric tons	145,150	-
, Tcal	786	703
Travel speed, km/hr	29.6	37.0
23,150 kilometres, round-trip:		
Round-trip travel, days	32.6	26.0
Loading/unloading/bad weather, days	5.0	5.0
Total trip, days	37.6	31.0
Round-trips per year	9.7	11.8
Energy delivered per tanker per year, Tcal	7,624	8,295
Project sizes & tanker requirements:		
Delivered product, metric tons/day	22,680	-
, Mnm^3/day	-	13.40
, Tcal/year	44,856 (3)	46,368 (3)
Tankers required	6	6

---

(1) Equivalent to 9,500 kcal/nm^3 (gas)

(2) 1 m^3 of LNG liquid = about 625 m^3 of natural gas

(3) Practically equivalent project sizes since regasification of LNG consumes some energy for vaporizing and for pumping (see LNG Regasification section). Methanol sold as liquid fuel with no regasification.

As can be seen, the overall methanol project requires considerably less capital, about $225,000,000 less. However, more capital must be invested overseas and this may raise considerations of political sensitivity.

As will be seen later in this section, the methanol project has a thermal efficiency of about 55%, compared with the LNG project's thermal efficiency of about 89% (see previous sections). Thus, the methanol project will consume about 60% more raw natural gas than will the LNG project to produce the same amount of energy. The cost of raw natural gas therefore becomes significant in the relative economics of the alternative projects. When all the factors of capital, operating and raw gas costs are considered, it has been estimated that the comparative end-product fuel values, delivered at the market terminal, are:

Round-trip distance, nautical miles	$/million Btu's Methanol	$/million Btu's LNG
24,000	1.10	1.60
16,000	1.05	1.30
8,000	0.98	0.98

(Based on a raw natural gas price of 11¢/million Btu's)

As mentioned above, the break-even point in competitiveness between the alternative projects is at a round-trip distance of about 8,000 nautical miles. At a round-trip distance of 16,000 miles (Indonesia to West Coast, U.S.A.) or of 24,000 miles (Persian Gulf to U.S.A.), the methanol project seems to be distinctly more economical.

However, these economics apply only if the methanol market can consume 25,000 tons/day for direct burning as a liquid. If the delivered methanol must be reconverted to pipeline SNG for distribution, the additional cost has been estimated as 20-25¢/million Btu's. In that event, it would appear that only the Persian Gulf-to-U.S.A. route shows an economic advantage for a methanol project over an LNG project.

TYPICAL PROCESS DESIGN FOR METHANOL SYNTHESIS

The synthesis of methanol from natural gas is a two-step process. First, the natural gas is 'reformed' to produce a gas containing primarily carbon monoxide and hydrogen. The gas is then converted to methanol in a 'methanol synthesis' unit, and the methanol is distilled to remove and recover water. The two chemical reactions can be written as:

$$CH_4 + H_2O \rightarrow CO + 3H_2 \rightarrow CH_3OH + H_2$$

As can be seen, the conversion produces excess hydrogen (H_2) as well as methanol (CH_3OH). Rather than waste this hydrogen, a supply of carbon dioxide (CO_2) can be reacted with the excess hydrogen to make additional methanol:

$$1/3\ CO_2 + H_2 \rightarrow 1/3\ CH_3OH + 1/3\ H_2O$$

The resulting overall reaction of converting methane plus carbon dioxide plus steam into methanol can be written as:

$$CH_4 + 1/3\ CO_2 + 2/3\ H_2O \rightarrow 4/3\ CH_3OH$$

Figure 7 presents a schematic flow sheet for converting natural gas into methanol, which involves five process steps:

> This description of a methanol plant is based on literature studies and estimates, and is not universally applicable. It is intended merely to illustrate the process involved.

Reforming -- Natural gas, CO_2 and steam are reformed, at about 200-300 psi pressure and 1500-1600°F, to yield the synthesis feed gas.

Compression -- The synthesis feed gas is compressed to the pressure required in the methanol synthesis unit.

Methanol Synthesis -- The synthesis gas is catalytically converted to methanol and water. Three process technologies are available for this synthesis:

(a) The ICI process (England)
(b) The Lurgi process (Germany)
(c) The Vulcan process (U.S.A.)

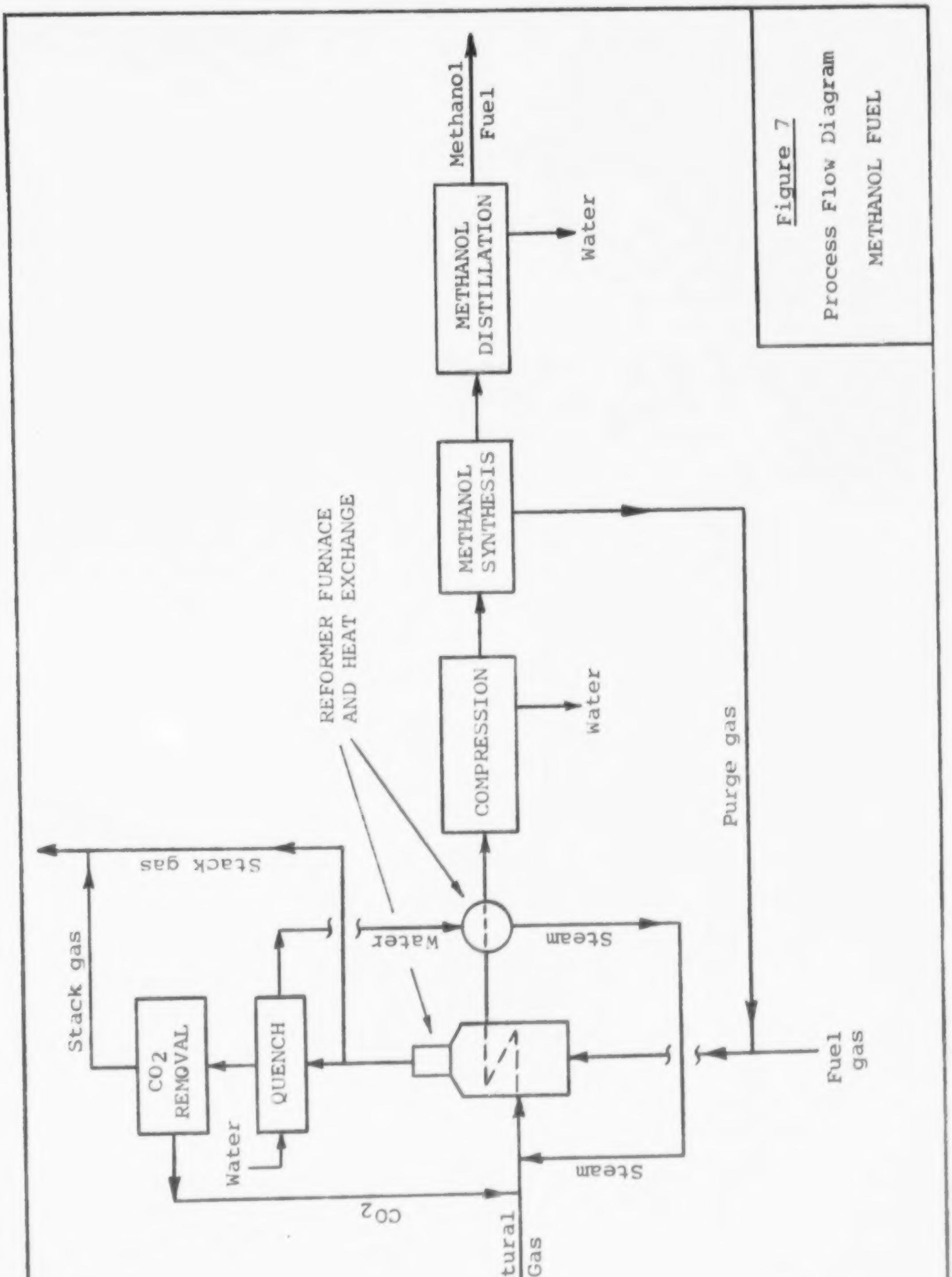
Natural Gas
CO2
Water
CO2 REMOVAL
Stack gas
QUENCH
Stack gas
Water
REFORMER FURNACE AND HEAT EXCHANGE
Steam
Steam
Fuel gas
Purge gas
COMPRESSION
Water
METHANOL SYNTHESIS
METHANOL DISTILLATION
Water
Methanol Fuel

Figure 7

Process Flow Diagram

METHANOL FUEL

These processes vary from 'low-pressure' to 'high-pressure' over a range of 750 to 4500 psi pressure, and a temperature range of 400-700°F.

Distillation -- The methanol product is boiled to remove and recover water. The product methanol is then stored for subsequent loading aboard tankers.

Stack Gas Quench and CO_2 Removal -- A portion of the stack gases from the reformer furnace is cooled by quenching with water. The cooled gases are then processed to remove and recover CO_2 as required in the production of methanol.

If we consider the flow scheme in Figure 7 as a single 'module', the largest such module under consideration today would produce about 5,000-7,500 tons/day of methanol. Thus, a 25,000 ton/day methanol project would require 4 to 5 parallel modules, each as shown in Figure 7.

THERMAL EFFICIENCY

The overall thermal efficiency of a plant producing 25,000 tons/day of methanol from natural gas can be approximated as:

	10^9 Btu/day (HHV)
Natural gas feed	530
Fuel gas to reformer	340
Fuel gas to electrical generation	15
TOTAL INPUTS	885
Product CH_3OH	488

Overall thermal efficiency = (444/885)(100) = 55%

The overall project efficiency is, in fact, lower yet because of the tanker fuel requirements. When we compare this very low efficiency with the 89% thermal efficiency of an LNG project, it becomes obvious that the methanol project involves a 'trade-off' of natural gas conservation versus lower cost storage and transportation. From the environmental viewpoint, can we afford a 55% efficiency versus an 89% efficiency in order to conserve capital?

OVERALL MATERIAL BALANCE

The material balance for the chemical processes involved can be approximated as:

	MM SCFD	Tons/day
Natural gas feed	525	11,100
Water (consumed chemically)	-	7,000
CO_2		8,600
TOTAL INPUTS		26,700
Product CH_3OH		25,000
Purge gas		1,700
TOTAL OUTPUTS		26,700

The material balance for the fuel gas combustion and CO_2 recovery can be expressed as:

	Tons/day
Reformer fuel gas	7,100
Purge gas to reformer fuel	1,700
Fuel gas to electrical generation	300
	9,100
Combustion air (w/10% excess)	172,800
TOTAL INPUTS	181,900
CO_2	8,600
Stack gases	173,300
TOTAL OUTPUTS	181,900

The overall material balance is the combination of the above two balances:

	Tons/day
Natural gas feed	11,100
Water (consumed chemically)	7,000
Fuel gases	7,400
Combustion air	172,800
TOTAL INPUTS	198,300
Product CH_3OH	25,000
Stack gases	173,300
TOTAL OUTPUTS	198,300

WATER BALANCE

Water is required in the process plant:

- as boiler feedwater to produce steam for reaction, for driving compressors, and for heat in the methanol distillation.
- as makeup to the cooling water system.
- for the usual potable and sanitation requirements.

Most of the steam used for compressor drives and for distillation heat can be recovered and reused as boiler feedwater. Hence, only the boiler and cooling water system blowdowns must be made up by fresh water.

It can be assumed that the excess reaction steam, when removed in the compression and distillation sections (see Figure 7), can be reused as boiler feedwater after some nominal treating.

Thus, there are only four ultimate consumptions of water which must be supplied by fresh water:

- Boiler blowdown
- Cooling water blowdown
- Chemical consumption of reaction steam
- Potable and sanitation uses

Chemical consumption of steam (i.e. water) amounts to 0.28 tons per ton of product, as shown in the overall material balance. The total use of fresh water for makeups and chemical consumption has been estimated as ranging from 1.5 to 2.5 tons per ton of product. For a 25,000 ton/day methanol plant, that would amount to 6,250-10,400 gpm -- a very large amount. The breakdown is approximated as:

	tons/ton product	gpm*
Chemical consumption	0.3	1,250
Cooling water makeup	1.0 - 2.0	4,150 - 8,300
Boiler feed makeup	0.2	850
	1.5 - 2.5	6,250 - 10,400

* For a 25,000 ton/day methanol plant, and neglecting the relatively insignificant amount of potable/sanitation water.

Obviously, if once-through seawater cooling is used rather than an evaporative cooling tower system, then the fresh water consumption might be as low as 0.5 ton per ton of product (2,000 gpm). Alternatively, extensive use of air-cooling would decrease the fresh water needs.

It might be interesting at this point to compare the fresh water usages of the various chemical conversion processes discussed in this report:

	Product Energy 10^9 Btu/day	Fresh Water gpm
Naphtha SNG	100	380$^{(1)}$
Coal SNG	310	5,100$^{(2)}$
Methanol Fuel	488	8,325$^{(3)}$
Oil SNG	1,733	12,620$^{(2)}$

(1) All air-cooling
(2) Extensive air-cooling
(3) Average of 6,250-10,400 gpm

This tabulation shows that the coal SNG and methanol plants are by far the largest users of fresh water per unit of product energy.

SUMMARY OF ENVIRONMENTAL FACTORS

Since all the fuel burned in the plant is natural gas, which must be purified prior to use as feedstock, there will be no stack gas emission problems of any kind. If the natural gas purification plant is included in the complex, it can be assumed that it includes a sulfur removal and recovery unit.

The major environmental consideration is the fresh water usage. The quantity required is very large, unless a once-through seawater cooling system is used.

The process effluents of excess reaction steam as condensed water (see Figure 7) must be treated and reused as boiler feedwater.

Socio-economic/socio-political factors are probably the most significant, i.e., can a process which is only 55% efficient in its conservation of energy be justified by a savings in capital investment, most of which must be invested in an overseas nation?

The other factors (noise levels, emergency flares, and construction personnel) will be very similar to those for the SNG plants discussed in previous sections.

ALTERNATIVE CONFIGURATIONS

This section has been devoted to the production and transportation of methanol from overseas sources of natural gas.

Considerable interest has also been evidenced in the production of methanol from coal. As discussed in Section V herein, the crude gas obtained from coal gasification is rich in carbon monoxide, carbon dioxide and hydrogen. This crude synthesis gas could be converted into methanol. In this case, the coal gasification process offers a number of opportunities to improve the thermal efficiency of the methanol synthesis. Therefore, methanol produced from coal may be more attractive than methanol produced from natural gas -- at least from the viewpoint of basic energy conservation.

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SECTION X

NATURAL GAS PIPELINES

Natural gas pipeline systems include many facilities other than the pipe through which the gas is transported. An overall system begins with the wellhead and gas treating facilities at the producing wells, and includes the compressor stations which move the treated gas through a pipeline network to the end-use market.

TYPICAL GAS PIPELINE SYSTEM

Figure 8 is a schematic flow diagram for a complete gas pipeline system. It includes three basic functional components:

-- The Wellhead Facilities:

The raw gas from a group of gas wells is piped to nearby wellhead separator stations which separate and remove water and associated oil from the gas.

-- The Field Gas Treating Plant:

The gas from a number of wellhead separator stations is then gathered and piped to a nearby field gas treating plant. As discussed earlier in this report (see Section II), the gas treating plant removes CO_2 and H_2S from the raw natural gas. In almost all cases, the H_2S will be converted to by-product sulfur.

The gas treating plant also dries the gas to meet pipeline transmission dewpoint specifications. (Although not shown in Figure 8, the treating plant would remove and recover by-product natural gas liquids (NGL) if they were present in sufficient quantities to make their recovery economically attractive). Finally, the treating plant provides treated gas for the local fuel needs of the field facilities.

-- The Gas Pipeline and Compressor Stations:

The treated natural gas then enters the pipeline transmission system for transport to market terminals. Large compressors are stationed along the pipeline (perhaps 100-200 miles apart) and are used to move the gas through the line.

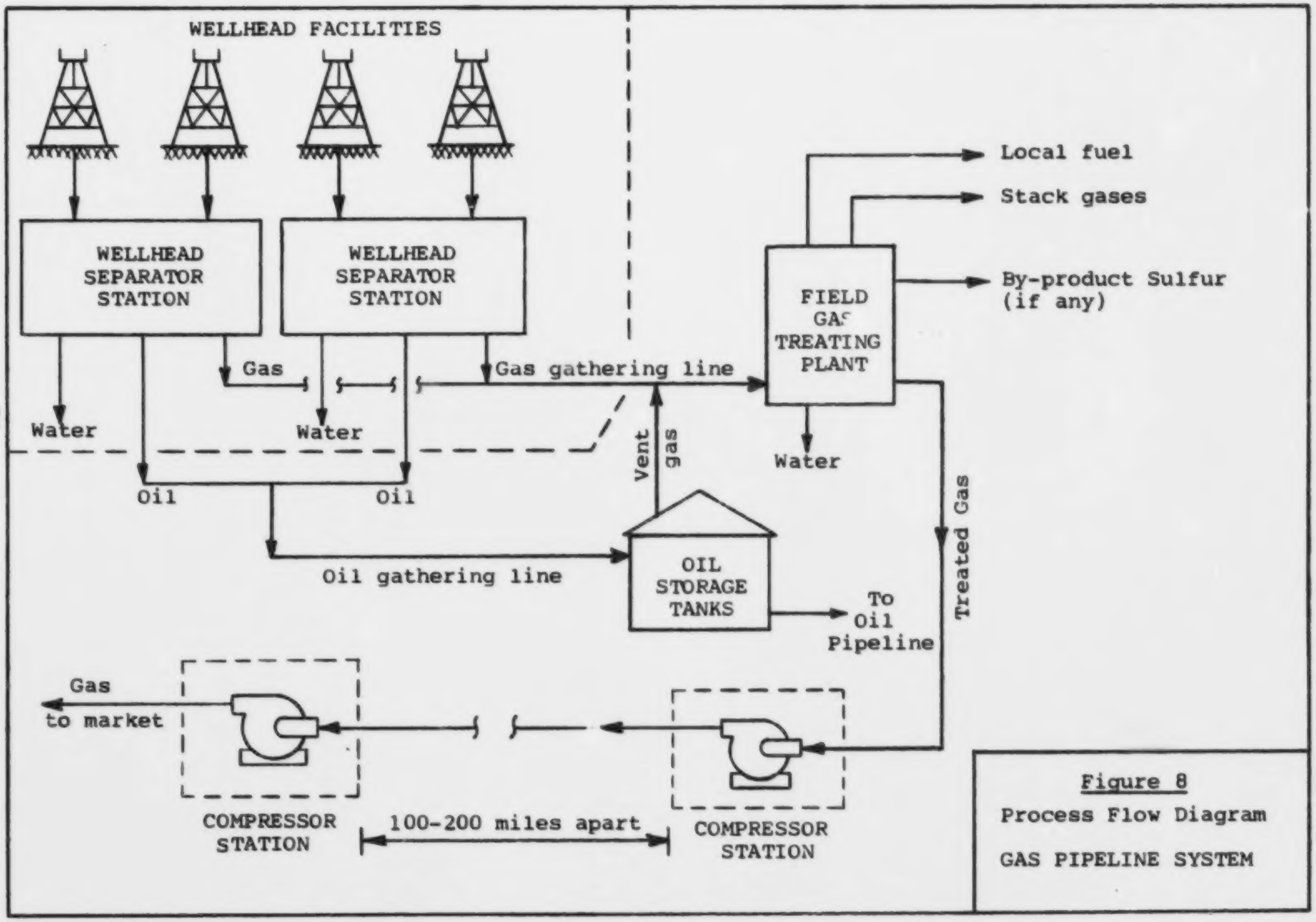
WELLHEAD FACILITIES
WELLHEAD SEPARATOR STATION
WELLHEAD SEPARATOR STATION
Gas
Gas gathering line
Water
Water
Oil
Oil
Oil gathering line
Vent gas
OIL STORAGE TANKS
To Oil Pipeline
FIELD GAS TREATING PLANT
Local fuel
Stack gases
By-product Sulfur (if any)
Water
Treated Gas
Gas to market
COMPRESSOR STATION
100-200 miles apart
COMPRESSOR STATION

Figure 8

Process Flow Diagram

GAS PIPELINE SYSTEM

Two types of compressors are used in pipelines. The oldest type is the reciprocating-piston compressor using a gas-fired engine as the motive driver. The more recent type is the centrifugal compressor using a gas-fired turbine as the motive driver. The gas-turbine centrifugals are rapidly gaining in popularity, although reciprocating units are still being used.

The purpose of discussing an overall system in Figure 8 is to emphasize that a pipeline project may include much more than the pipe through which the treated gas moves. All pipeline projects, however, will not necessarily include all of the components shown in Figure 8. Some wells do not produce associated oil, and some raw gases may not contain much CO_2 or H_2S or significant amounts of NGL. Some projects may only involve additions to an existing system, or new lateral transmission lines to some additional market terminals. In other words, some projects may include only minimal field facilities while others may not include such facilities at all.

GAS TRANSMISSION PIPELINE COSTS

Tables 11 and 12 present some detailed statistics and cost data regarding gas pipelines. In brief, current costs for the pipe itself have been estimated to average about $6,800 per mile per inch of pipe diameter. Thus, a 1,000 mile pipeline of 36-inch diameter would currently cost about $245,000,000. This includes rights-of-way, materials and labor for a line installed on land. It does not include such unusual conditions as may be encountered by an Arctic pipeline.

The cost of undersea lines installed offshore might be 3 to 6 times as much as a pipeline on land.

The current cost of compressor stations (including land, buildings, materials and labor) has been estimated to average about $305 per horsepower for centrifugal compressors and about $565 per horsepower for reciprocating compressors.

TABLE 11E -- GAS PIPELINE COSTS - ENGLISH UNITS

Basis:
(a) Costs based on 1969 completions.
(b) All lines are on land (no offshore lines).
(c) Lines include main transmission pipelines as well as lateral distribution lines.
(d) Costs include rights-of-way, materials, labor and other costs.
(e) Costs do not include compressor stations, field gas gathering systems, or field gas treating plants.
(f) Costs do not include unusual conditions (such as for Arctic pipelines).

Line Diameter (inches)	Cost per mile ($)	Cost per Inch-Mile ($/inch-mile)
6	36,000	6,000
8	41,000	5,125
10	50,200	5,020
12	60,000	5,000
16	78,400	4,900
20	101,800	5,090
24	171,840	7,160
30	175,200	5,840
36	216,720	6,020
42	267,540	6,370
Range (1969 completions)		5,000 - 7,160
Average (1969 completions)		5,650
Average (1974 completions)(1)		6,780

---

(1) Based on 20% inflation during 5 year period of 1969-1974

TABLE 11M -- GAS PIPELINE COSTS- METRIC UNITS

Basis:
(a) Costs based on 1969 completions
(b) All lines are on land (no offshore lines)
(c) Lines include main transmission pipelines as well as lateral distribution lines
(d) Costs include rights-of way, materials, labor and other costs
(e) Costs do not include compressor stations, field gas gathering systems, or field gas treating plants
(f) Costs do not include unusual conditions (such as for Arctic pipelines)

Line Diameter (cm)	Cost per km ($)	Cost per cm/km ($/cm/km)
15.2	22,370	1,470
20.3	25,480	1,255
25.4	31,200	1,230
30.5	37,290	1,225
40.6	48,730	1,200
50.8	63,270	1,245
61.0	106,800	1,750
76.2	108,890	1,430
91.4	134,690	1,475
106.7	166,280	1,560
Range (1969 completions)		1,200 - 1,750
Average (1969 completions)		1,385
Average (1974 completions)(1)		1,660

---

(1) Based on 20% inflation during 5 year period of 1969-1974

TABLE 12M -- SOME GAS PIPELINE STATISTICS - ENGLISH UNITS

1 -- Total daily gas transmission = 1.69×10^9 nm^3/day(1)

2 -- Total miles of pipeline:

gas transmission = 302,490 km(1)
field gas gathering = 94,930 km(1)

3 -- Total transmission compressor stations(1)

= 1,030 stations
= 8,453 MW
= average of 8.2 MW per station(2)
= average of 1 station per 294 km of transmission pipeline

4 -- Total field gas processing compressor stations(1)

= 725 stations
= 1,491 MW
= average of 2.1 MW per station(2)

5 -- New transmission compressors (completed in 1969):

	Range	Average	
Centrifugal (gas-turbine):			
KW	746-9,321	4,937	(4)
Cost, \$/KW(3)	232-654	355	
Reciprocating (gas-engine):			
KW	373-6,711	3,803	(4)
Cost, \$/KW(3)	450-762	660	

---

(1) In U.S.A. at end of 1972

(2) May be more than 1 compressor per station

(3) Cost includes land, buildings, equipment and labor

(4) Assuming 15% escalation during 5 year period of 1969-1974, current costs would be 408 \$/KW for centrifugal and 759\$/KW for reciprocating

TABLE 12E -- SOME GAS PIPELINE STATISTICS - METRIC UNITS

1 -- Total daily gas transmission = 63 x 10^9 SCFD(1)

2 -- Total miles of pipeline:

gas transmission = 188,000 miles(1)
field gas gathering = 59,000 miles(1)

3 -- Total transmission compressor stations(1)

= 1,030 stations
= 11,335,000 horsepower
= average of 11,000 horsepower per station(2)
= average of 1 station per 183 miles of transmission pipeline

4 -- Total field gas processing compressor stations(1)

= 725 stations
= 2,000,000 horsepower
= average of 2,760 horsepower per station(2)

5 -- New transmission compressors (completed in 1969):

	Range	Average
Centrifugal (gas-turbine drive):		
Horsepower	1,000-12,500	6,620
Cost, $/HP(3)	173-488	265(4)
Reciprocating (gas-engine drive):		
Horsepower	500-9,000	5,100
Cost, $/HP(3)	336-568	492(4)

---

(1) In U.S.A. at end of 1972

(2) May be more than 1 compressor per station

(3) Cost includes land, buildings, equipment and labor

(4) Assuming 15% escalation during 5 year period of 1969-1974, current costs would be 305 $/HP for centrifugal and 565 $/HP for reciprocating.

Based on 1972 U.S. statistics, there is an average of one compressor station of 11,000 horsepower per each 183 miles of gas transmission pipeline in the country (see Table 12). Thus, the average 1,000-mile pipeline would have about 5 or 6 compressor stations with a total horsepower of about 60,100. Assuming centrifugal compressors were used, these would currently cost about $18 million. Therefore, an average 1,000-mile transmission pipeline of 36-inch diameter would have a total cost of about:

Pipeline	$245,000,000
Compressors	18,000,000
	$263,000,000

Costs of field gas gathering and treating are too specific and too variable to permit any generalized cost estimates.

COMPARISON TO SNG, LNG AND METHANOL PROJECTS

At the end of 1972 (see Table 12), the U.S. had a total of 188,000 miles of gas transmission pipelines with 11,335,000 compressor horsepower. The network transported 63 billion cubic feet of gas per day. Relating this to the previous sections on SNG, LNG and methanol projects:

	% of 1972 Pipeline Transmission
250 MM SCFD of SNG	0.4
500 MM SCFD of LNG	0.8
25,000 tons/day of methanol	0.8

Obviously, a great many of these alternative fuel supply projects will be required to alleviate current and projected energy shortages.

ENVIRONMENTAL FACTORS

The environmental factors relating to wellhead and field gas treating facilities are too specific and too diverse to attempt any generalized statements.

As for the pipeline and compressor stations, the major environmental impacts will be those associated with land disturbances along the rights-of-way (ROW). Land disturbance factors will vary widely, depending upon the characteristics of the area through which the pipeline is routed. In some locations, the pipelines may require ROW fencing and access roads, which will cause other environmental effects.

The noise levels of large compressors are very high. However, in most instances, compressor stations are in remote, non-urban areas and enclosed. Therefore, noise level impacts on communities should be minimal.

Usually, compressor drivers are fueled by the clean-burning pipeline gas. Therefore, SO_2 and NO_x emissions from the driver stack gases should be low. As a good approximation for gas-turbines burning pipeline gas:

	Per Horsepower
Fuel burned	11,000 Btu/hr
Wet Flue Gas rate	340 SCF/hr
NO_x emissions	0.0033 lbs/hr
SO_2 emissions	0.000015 lbs/hr

Thus, a 10,000-horsepower compressor station will result in the following:

Fuel burned:	
Btu/hr	110,000,000
SCF/hr	110,000
Wet Flue Gas, SCF/hr	3,400,000
NO_x emissions:	
lbs/hr	33
lbs/MM Btu	0.3
SO_2 emissions:	
lbs/hr	0.15
lbs/MM Btu	0.0014

The above data are based on a fuel gas containing about 0.5 grains H_2S/100 SCF and on the gas-turbine using about 225% excess air

Socio-economic impacts caused by construction personnel will depend upon the size and duration of the project and the number of workers required.

O'Donnel, J. P. Pipeline Growth Remains Slow. Oil and Gas Journal 71:24 112-119, June 1973.

O'Donnell, J. P. 13th Annual Study of Pipeline Installations and Equipment Costs. Oil and Gas Journal 68:31 99-120, August 1970.

Trans-Alaska Report: Alyeska North Slope. Oil and Gas Journal 72:11 52-110, March 1974.

SECTION XI

LIQUID FUELS FROM OIL SHALE

Geologically speaking, oil shale is not a shale and it contains virtually no oil. It is a sedimentary rock containing a solid organic material called 'kerogen'. When heated, the kerogen yields substantial amounts of hydrocarbon crude oil and gas -- ranging typically from 10 to 60 gallons of crude oil per ton of shale. Aeons ago, oil shale had a beginning similar to conventional petroleum crude oil, when organic matter was deposited in large and ancient lakes. However, the oil shale deposits were not subjected to the heat and pressure required to form petroleum. Instead, the organic matter was transformed into the solid hydrocarbon kerogen and locked into a marlstone matrix. The geological term for our Western oil shale is 'kerogenous marlstone'. A typical oil shale contains about 15 wt% kerogen and 85 wt% of carbonates, feldspars, quartz and clays:

Kerogen content	15 wt%	
Kerogen composition:		wt%
Carbon		80.5
Hydrogen		10.3
Nitrogen		2.4
Sulfur		1.0
Oxygen		5.8
		100.0
Mineral content	85 wt%	
Carbonates		48.0
Feldspars		21.0
Quartz		15.0
Clays		13.0
Analcite and Pyrite		3.0
		100.0

Oil shale is found on every continent throughout the world. Reserves of oil shale are usually expressed in terms of the barrels of oil contained in the shale deposits, as determined by a standard laboratory analysis. Over 15 nations around the world have extensive shale oil reserves, the largest of which are:

	In-place Reserves
Brazil	342 billion barrels
U.S.A.	2,000 billion barrels
Zaire	103 billion barrels

In the United States, shale deposits occur in a great many states, the most prominent being in these areas:

Colorado-Utah-Wyoming
Montana
California
Alaska
Texas-Oklahoma-Arkansas
Eastern U.S.

From the viewpoint of reserves and recoverability, the best oil shale deposits for development in the U.S. are the Green River Formation deposits in Colorado, Utah and Wyoming. Of these, the Piceance Basin deposits are considered to be of prime importance. As shown in Figure 9, the Green River Formation encompasses about 17,000 square miles and has an in-place oil potential that has been estimated to be at least 2,000 billion barrels. At our current national oil consumption rate of about 18 million barrels a day, the Green River Formation represents about a 300 year supply of in-place oil. It has also been estimated that about 85 billion barrels of the total in-place reserves could be recovered with today's mining technology. Thus, about a 12 year national supply of oil could be recovered without requiring any technological break-throughs in mining methods. To put this in perspective, it is almost as much oil as produced in the entire U.S. since the Civil War (1860). And we can assume that advances in mining technology will progressively make more of our in-place reserves recoverable, including those in the other U.S. oil shale deposits.

Commercial shale oil production has been practiced for many years. It began as early as 1840 in France and Scotland, and subsequently in other countries. Through the years, most of these operations ceased

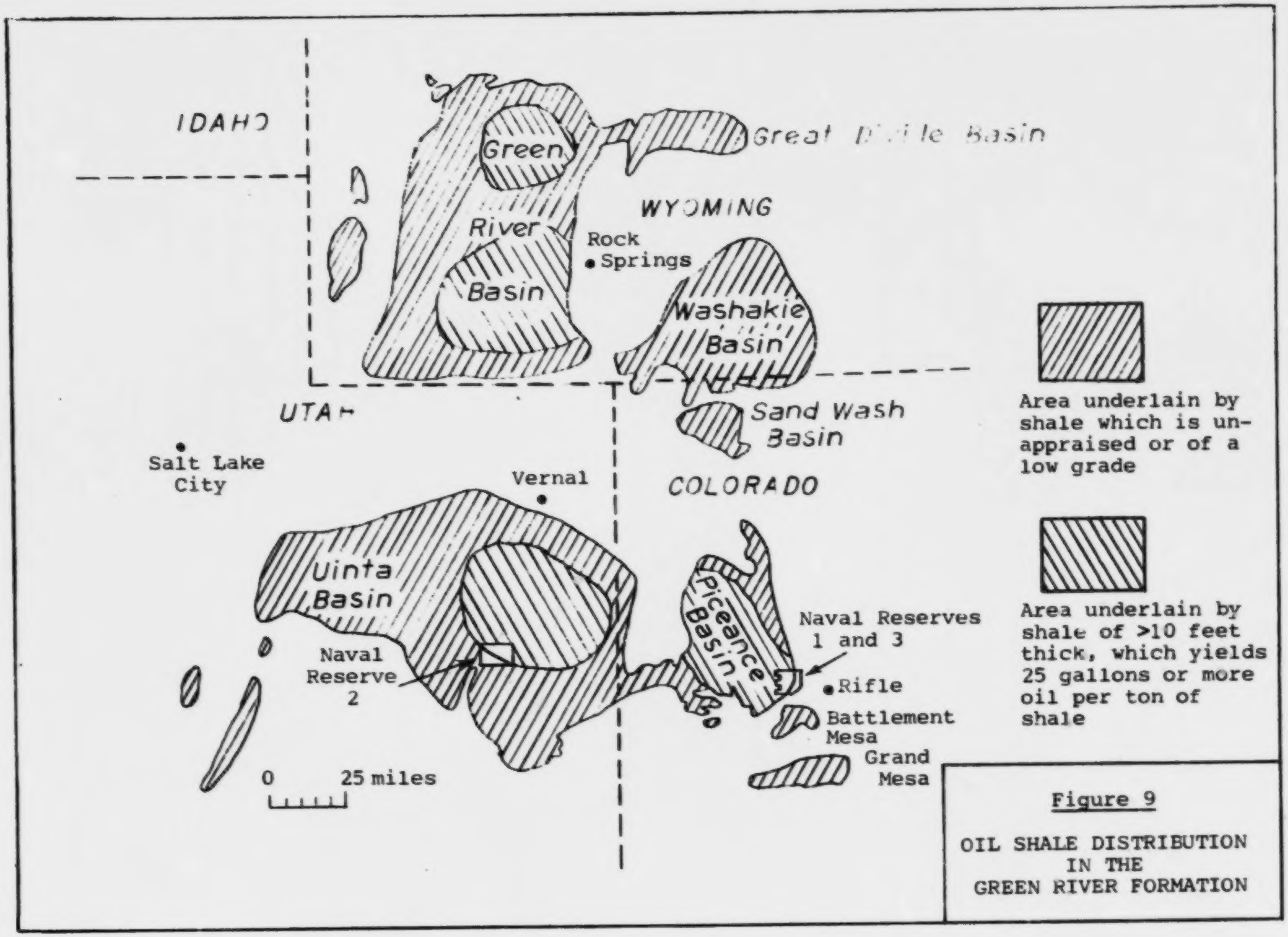
IDAHO
Great Divide Basin
Green
River
Basin
WYOMING
Rock Springs
Washakie Basin
Sand Wash Basin
UTAH
Salt Lake City
Vernal
COLORADO
Uinta Basin
Piceance Basin
Naval Reserves 1 and 3
Rifle
Battlement Mesa
Grand Mesa
Naval Reserve 2
0
25 miles
Area underlain by shale which is unappraised or of a low grade
Area underlain by shale of >10 feet thick, which yields 25 gallons or more oil per ton of shale

Figure 9

OIL SHALE DISTRIBUTION IN THE GREEN RIVER FORMATION when petroleum crude oil became available. China and Estonia (now in the U.S.S.R.) have the only commercial shale oil production operations today. China produces about a 100,000 barrels a day, principally in Manchuria. Estonia has been producing shale oil for 50 years and their current output is perhaps 50,000 barrels a day of oil plus about 100 MM SCFD of low-Btu fuel gas. In Brazil, a semi-commercial prototype plant designed to produce 1,000 barrels a day of oil and over 1 MM SCFD of gas is currently in startup. For contrast, the forecasts for near-term U.S. shale oil production in the Green River Formation area range from 1 to 4 million barrels a day. At that rate, the currently recoverable 85 billion barrels of oil in that area would last from 60 to 240 years.

As a broad generality, the raw shale oil that can be extracted from Western oil shales has the following characteristics:

Gravity, °API	18-28
, lbs/gal	7.882-7.387
Pour point, °F	30-90
Sulfur, wt %	0.6-0.8
Nitrogen, wt %	1.6-2.2
Distillates, vol % :	
Naphtha and lighter	18-24
Diesel oil	24-17
Fuel oil	34-33
Residuum	24-26

Assuming an oil shale containing 15 wt % of kerogen (of which about 90% is recoverable hydrocarbon with a gravity of about 23°API), the yield of raw shale oil would be 35 gallons per ton of shale. Thus, a shale oil extraction project producing 50,000 BSD of raw shale oil would require 60,000 tons per day of oil shale feedstock. If the near-term U.S. shale oil production should reach 4 million barrels a day, then the mining of oil shale feedstock might be expected to approach 5 million tons per day.

CURRENT DEVELOPMENT PROJECTS

During the last 15-20 years, the U.S. Bureau of Mines and various oil companies have investigated many processes for extracting shale oil from our Western oil shales. Most of these processes fall into two categories -- (1) retorting of mined shale and (2) in-situ combustion (i.e. underground in-place retorting). Recently, a University of California research team has begun a study of biochemical leaching as another approach.

At this time, the most advanced retorting processes from the viewpoint of testing are:
-- The Oil Shale Corporation's TOSCO II process
-- Union Oil Company's Steam Gas Recirculation (SGR) process
-- Development Engineering Incorporated's Parahoe process
-- The Lurgi-Ruhrgas process
The Institute of Gas Technology and the Bureau of Mines have other oil shale processes in development.

Currently, the major oil shale projects actively underway include the following:

-- The ARCO Colony project is designing a 50,000 BSD shale oil plant at Parachute Creek, Colorado using the TOSCO II process. Although a construction contract for the plant had been awarded, inflation has raised the plant cost estimates and resulted in shutting down the project for the time being.

-- Union Oil has tested a 3 ton/day SGR unit at Parachute Creek and is proceeding with a 1,500 ton/day pilot plant. Union Oil ultimately plans a 50,000 BSD shale oil project.

-- Standard Oil of Ohio (SOHIO) heads a 17-company consortium in operating a $7,500,000 demonstration project in the Bureau of Mines facility at Anvil Points, Colorado. The demonstration project is using a 10½ foot diameter Parahoe retort.

-- Garret Research and Development, a subsidiary of Occidental Petroleum, is testing and developing a modified in-situ process.

-- Superior Oil plans to test and develop a retorting process to recover shale oil as well as nacohlite (soda ash) and dawsonite (an aluminum-bearing mineral).

SHALE OIL PRODUCTION

An overall block flow diagram for a shale oil production project is depicted in Figure 10. Based on above-ground retorting, the individual steps in the project would include:

Mining -- Either surface or underground mining of the oil shale could be used. However, the Green River Formation deposits are not particularly well suited for the various surface mining techniques -- strip mining, open pit mining, or the so-called 'glory hole' mining. In all probability, the room and pillar method of underground mining (with access by adit, slope or shaft tunnels) will be used in the Green River Formation deposits.

Raw shale stockpile -- Primary crushing of the mined shale will be at the mine portal, and the coarsely crushed shale will be conveyed to the raw shale stockpile. For a 50,000 BSD shale oil project, involving 60,000 tons/day of mined shale, the coarse shale stockpile will contain about 500,000 tons which will cover about seven acres to a height of about 200 feet.

Crushing -- The shale is then conveyed to the final crushers which will produce a fine shale of less than ½ inch in size. Conveyors will carry the fine shale to storage silos containing about 15,000 tons (about a 6-hour supply).

Retorting -- The retorts, or pyrolysis units, will use heat to vaporize and remove the kerogen from the shale. Figure 11 is a block flow diagram of the TOSCO II retorting process. As shown in Figure 11, heat is transferred into the shale by solid-to-solid heat exchange between the shale and hot ceramic balls. The balls are heated in a vertical ball heater and then fed into the retort to mix with preheated shale. The resulting pyrolysis temperature is 900°F, which converts the kerogen into hydrocarbon oil vapors.

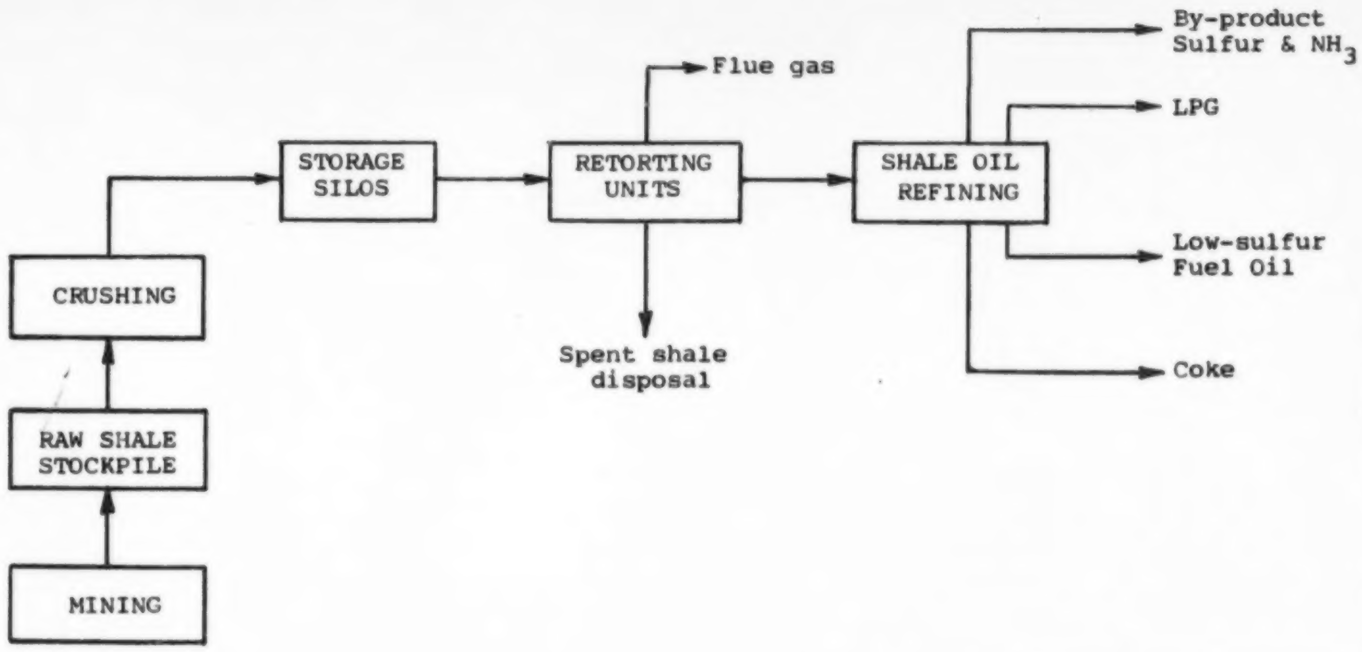
STORAGE SILOS
RETORTING UNITS
SHALE OIL REFINING
CRUSHING
RAW SHALE STOCKPILE
MINING
Flue gas
Spent shale disposal
By-product Sulfur & NH3
LPG
Low-sulfur Fuel Oil
Coke

Figure 10

Process Flow Diagram

SHALE OIL PRODUCTION

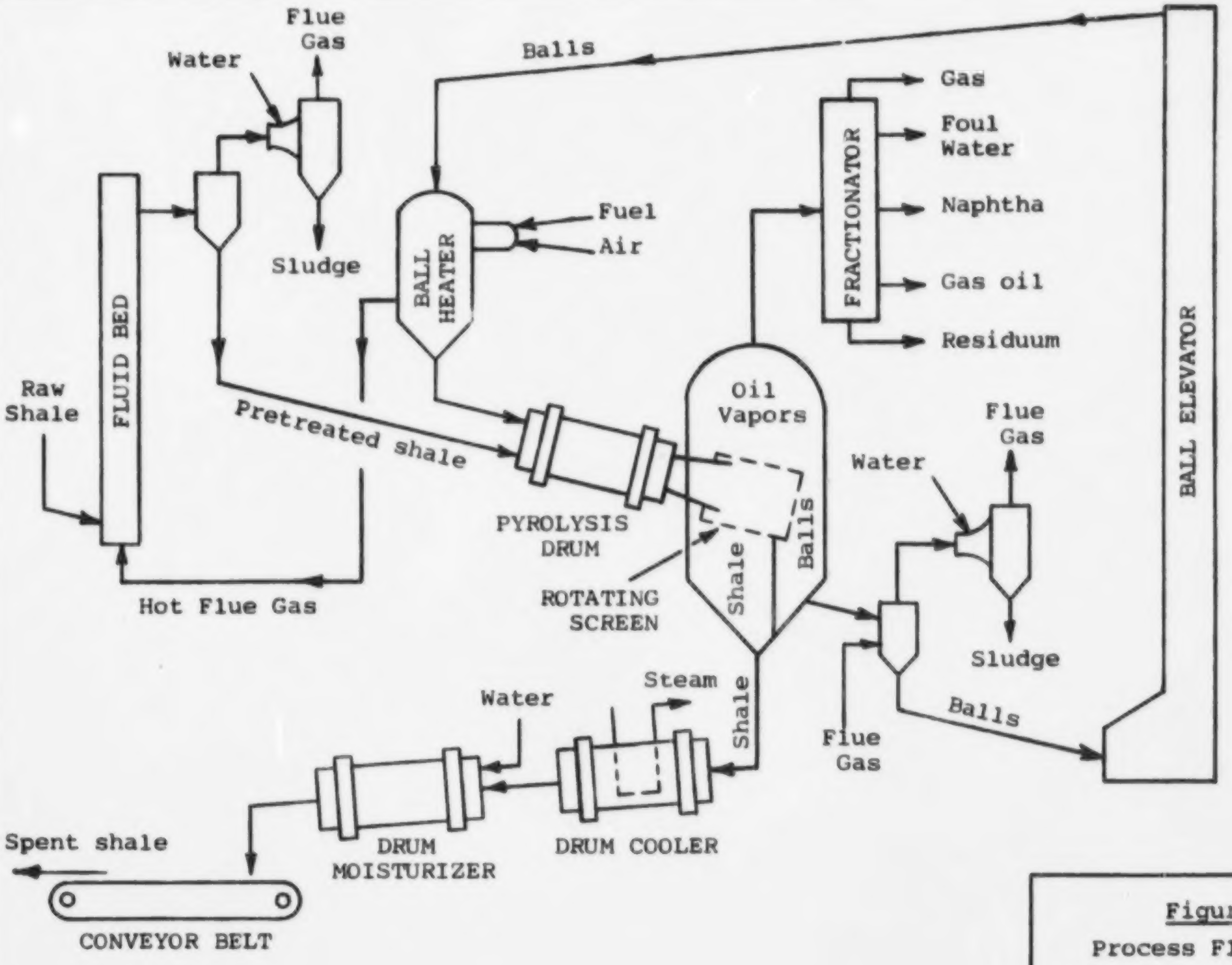
Flue Gas
Water
Balls
Gas
Foul Water
Fuel
Air
Sludge
BALL HEATER
FRACTIONATOR
Naphtha
FLUID BED
Gas oil
Residuum
Raw Shale
Oil Vapors
Pretreated shale
Flue Gas
BALL ELEVATOR
Water
PYROLYSIS DRUM
Balls
Shale
Hot Flue Gas
ROTATING SCREEN
Sludge
Steam
Water
Shale
Flue Gas
Balls
Spent shale
DRUM MOISTURIZER
DRUM COOLER
CONVEYOR BELT

Figure 11

Process Flow Diagram

TOSCO II RETORTING

The ceramic balls, spent shale and oil vapors are disengaged in a rotating screen (or 'trommel') within the retort. The spent shale is cooled, moisturized and conveyed to disposal. The balls are returned to the ball heater by a vertical-lift ball elevator. The oil vapors are fractionated and separated into the crude products of gas, naphtha, gas oil and residuum oil. Foul water, condensed in the fractionation, is sent to a stripper for removal of H_2S and ammonia. The stripped water is then used to moisturize the spent shale.

Shale oil refining -- The raw hydrocarbon products from the retorts will be upgraded in conventional oil refining units. These will include:

-- Gas recovery and acid gas removal
-- Synthesis of hydrogen for catalytic desulfurizers
-- Naphtha desulfurizing and gas oil desulfurizing
-- Coking to convert residuum oil into additional naphtha and gas oil
-- Strippers to remove H_2S and ammonia from sour waters
-- Sulfur recovery and ammonia recovery

The sales products (see Figure 10) will be LPG, low-sulfur fuel oil, petroleum coke, and by-product sulfur and ammonia. The project's internal fuel needs will be supplied by by-product fuel gas, butanes and some fuel oil.

Auxiliary services -- These will include a boiler plant to generate steam, a closed-loop evaporative cooling tower, water treatment, and other systems. In addition, about 85 MW of purchased electrical power will be needed during normal operation.

> This description of a shale oil production plant is based on a specific design and process at a specific site and it is not universally applicable. It is intended merely to illustrate the processes involved.

OVERALL PROJECT MATERIAL BALANCE

The overall material balance for a complete shale oil production project, such as is described above, can be summarized as follows:

	Tons/day(1)	BSD	wt %
INPUT:			
Raw shale	66,000	-	-
OUTPUTS:			
C_3 LPG	385	4,330	0.6
Low-sulfur fuel oil	7,200	47,000	10.9
Petroleum coke	800	-	1.2
By-product sulfur	157	-	0.2
By-product ammonia	135	-	0.2
Total sales products	8,677		13.1
Spent shale(2)	53,743	-	81.5
CO_2 vent stack(3)	355	-	0.5
Shale dust emissions(4)	10	-	negl.
Water of retorting(5)	900	-	1.4
Project fuel(6)	2,315	-	3.5
Total products	66,000		100.0

(1) Short tons (2) Excludes moisturizing water

(3) Only the carbon portion of the hydrogen plant CO_2 vent

(4) From vent stacks in the retorting units

(5) Water formed chemically during retorting of the shale

(6) 4,060 x 10^6 Btu/hr of fuel with a heating value of 21,025 Btu/lb. Equivalent to 15,700 BSD of a typical fuel oil with a heating value of 6.2 x 10^6 Btu per barrel.

This balance does not include combustion air for heaters and boilers, and the resultant stack gases. Nor does it include water usage and disposition.

OVERALL THERMAL EFFICIENCY

The total hydrocarbons obtained from the shale (the sales products plus the project fuel) amount to 10,992 tons/day as seen above. Assuming an average heating value of about 20,000 Btu/lb, the total hydrocarbons in the shale therefore contain 440 x 10^9 Btu/day of heating value.

Since the project consumes 97.5 x 10^9 Btu/day of fuel, the apparent thermal efficiency is 77.8% :

	Tons/day	10^9 Btu/day
Fuel gas	1,023	45.5
Fuel oil	832	32.5
Butanes	460	19.5
	2,315	97.5

The use of 85 MW of purchased elctricity requires the burning of 20 x 10^9 Btu/day of fuel for power generation. If this is included in the energy balance, the thermal efficiency is reduced to 73.3%.

OVERALL SULFUR BALANCE

The overall sulfur balance for the shale oil project will be:

	Tons/day as Sulfur	Actual form present
INPUT:		
Sulfur in raw shale(1)	528	sulfur compounds
OUTPUTS:		
Sulfur in LPG product	neglig.	-
Sulfur in fuel oil product(2)	36	organic sulfur
Sulfur in coke product(3)	4	organic sulfur
By-product sulfur	157	sulfur
Sulfur in spent shale(4)	323	sulfur compounds
SO_2 emissions from burning project fuel	7.6	SO_2
SO_2 emissions from sulfur recovery tail gas	0.4	SO_2
	528	

(1) Based on 0.8 wt % sulfur in the raw shale

(2) Based on 0.5 wt % sulfur in the low-sulfur fuel oil

(3) Based on 0.5 wt % sulfur in the coke

(4) Based on 0.6 wt % sulfur in the spent shale

(5) Based on 14 mol wt. fuel gas with 10 grains H_2S per 100 SCF, project fuel oil with 0.8 wt % sulfur, and butanes with 0.1 wt % sulfur. The quantities of gas, oil and butanes are as seen above.

The tail gas sulfur emission is about 0.25% of the sulfur fed to the recovery units. In other words, the sulfur recovery units are about 99.75% efficient.

STACK GASES

Table 13 is an itemized listing of the plant stacks with the quantity and composition of their stack gas effluents. The tabulation does not include the emergency flare stack, nor does it include other miscellaneous vents.

The total emissions of SO_2, NO_x and particulates, during normal operations, are summarized below:

	Tons/day
SO_2	16.0
NO_x	12.2
Particulates	13.0

The SO_2 emissions are based on the overall sulfur balance and on the project fuel consumption in combustion units (heaters, boilers, etc.). It was assumed that each combustion unit will use its prorata share of gas, butanes and oil based on each combustion unit's heat release. The overall emission of SO_2 from fuel combustion (15.2 tons/day) amounts to 0.31 lbs/MM Btu of heat release. This is well within the EPA's New Source Performance Standards (NSPS) for fuel burning --which are 1.2 and 0.8 lbs/MM Btu for coal and oil burning, respectively.

The estimated NO_x emissions (12.2 tons/day) are based on achieving 0.25 lbs NO_x/MM Btu, or less. This is consistent with the EPA's NSPS for fuel burning -- which are 0.2 and 0.3 lbs/MM Btu for gas and oil, respectively.

The estimated particulates emissions from combustion (4.9 tons/day) amount to 0.1 lbs of particulates/MM Btu. The EPA's NSPS for any fossil fuel is 0.1 lbs/MM Btu, and it was assumed that this standard could be met.

The particulates emissions from the spent shale moisturizer (3.1 tons/day), and from the crushing and conveying systems (5.0 tons/day)

TABLE 13E -- STACK GASES FROM SHALE OIL PRODUCTION- ENGLISH UNITS

	Project Fuel Heat Release	Total Stack Gases			Tons/day of Emissions		
	MM Btu/day	MM SCFD	Tons/day	°F	SO_2	NO_x	Partic.
Retorting flue gas:							
Ball heater	58,900	1,561(1)	57,660	130	9.2	7.3	3.0
Ball circulation	4,400	322(1)	11,890	140	0.7	0.6	0.2
Refining heaters	30,800	339(2)	12,520	500	4.8	3.9	1.5
Boilers	3,400	37(2)	1,370	500	0.5	0.4	0.2
	97,500				15.2(3)	12.2(4)	4.9(5)
Sulfur recovery tail gas	-	40	1,700	100	0.8(3)	nil	nil
Spent shale moisturizer	-	144	3,570	200	nil	nil	3.1
Crushing and conveying(6)	-	907	34,700	60	nil	nil	5.0
CO_2 vent	-	92	2,900	200	(45 wt% CO_2, 55 wt% H_2O)		
					16.0	12.2	13.0

---

(1) From venturi wet scrubbers.

(2) Based on 11,000 SCF of wet flue gas per 10^6 Btu of fuel heat release.

(3) Equivalent to 7.6 tons/day of elemental sulfur from fuel burning and 0.4 tons/day of elemental sulfur from tail gas (as shown in overall sulfur balance).

(4) Based on achieving 0.25 lbs NO_x/MM Btu of heat release. EPA standards are 0.2 and 0.3 lbs NO_x/MM Btu for firing gas and oil, respectively.

(5) Based on achieving 0.1 lbs particulates/MM Btu of heat release. EPA standard is 0.1 lbs/MM Btu for any fossil fuel.

(6) Dust collection system stacks at primary crusher, final crusher and shale fines storage silos.

TABLE 13M -- STACK GASES FROM SHALE OIL PRODUCTION - METRIC UNITS

	Project Fuel Heat Release	Total Stack Gases			Mg/day of Emissions		
	Gcal/day	Mnm^3/day	Mg/day*	°C	SO_2	NO_x	Partic.
Retorting flue gas:							
Ball heater	14,842	41.8(1)	52,309	54	8.3	6.6	2.7
Ball circulation	1,109	8.6(1)	10,787	60	0.6	0.5	0.2
Refining heaters	7,762	9.1(2)	11,358	260	4.4	3.5	1.4
Boilers	857	1.0(2)	1,243	260	0.5	0.4	0.2
	24,570				13.8(3)	11.0(4)	4.5(5)
Sulfur recovery tail gas	-	1.1	1,542	38	0.7(3)	nil	nil
Spent shale moisturizer	-	3.9	3,239	93	nil	nil	2.8
Crushing and conveying(6)	-	24.3	31,480	16	nil	nil	4.5
CO_2 vent	-	2.5	2,631	93	(45 wt% CO_2, 55 wt% H_2O)		
					14.5	11.0	11.8

---

(1) From venturi wet scrubbers

(2) Based on 1,169 nm^3 of wet flue gas per Gcal of fuel heat release.

(3) Equivalent to 6.9 metric tons/day of elemental sulfur from fuel burning and 0.36 metric tons/day of elemental sulfur from tail gas.

(4) Based on achieving 0.45 kg NO_x/Gcal of heat release. EPA standards are 0.36 and 0.54 kg NO_x/Gcal for firing gas and oil, respectively.

(5) Based on achieving 0.18 kg particulates/Gcal of heat release. EPA standard is 0.18 kg/Gcal for any fossil fuel.

(6) Dust collection system stacks at primary crusher, final crusher and shale fines storage silos.

* Mg/day is equivalent to metric tons/day

amount to 0.3 and 0.08 grains of particulates/SCF respectively. These are consistent with the control systems used for removal of dust particles in those services -- wet scrubbers for the moisturizer vent, and baghouses for the crushing and conveying.

The control of NO_x emissions to the levels discussed above (0.25 lbs/MM Btu) will require special combustion design features such as two-stage combustion, low usage of excess air, flue gas recirculation, and perhaps preferential use of fuel gas and butanes in certain services (rather than fuel oil).

OVERALL WATER BALANCE

The total project, including mining and spent shale disposal, will require 5,475 gpm of water. As shown in Table 14, the total fresh water requirement will be 4,970 gpm (about 8,000 acre-ft-year).

Almost all of the water supplied to the project will eventually return to the atmosphere. As seen in Table 14, 71.4% returns directly to the atmosphere as evaporation, and from vents and stacks. Another 25.9%, for revegetation and shale moisturizing, will eventually return to the atmosphere either directly or indirectly.

There will be no direct discharge of wastewaters to any natural waterway. Practically all of the effluent wastes from the project, after stripping for removal of H_2S and NH_3 and treatment for oil removal, are disposed of with the spent shale. Whether or not this will eventually contaminate underground aquifers, nearby streams, or surface run-off will depend upon a broad range of complex factors such as:

-- Whether the spent shale is disposed of by backfilling the mine, or by placement in a surface embankment. At least one of the major projects, under study at this time, is currently planning to use surface disposal.

TABLE 14 -- WATER REQUIREMENTS AND DISPOSITION

	gpm	m^3/hr	%
WATER REQUIREMENTS			
Fresh water:			
Dust suppression(1)	1,025	233	18.7
Revegetation	70	16	1.3
Boilers, cooling tower, utility	3,055	693	55.8
Shale moisturizing makeup	820	186	15.0
Total fresh water	4,970(2)	1,128	90.8
Other water:			
Natural shale surface water	75	17	1.4
Water of retorting(3)	150	34	2.7
Water of combustion(3)	280	64	5.1
Total water required	5,475	1,243	100.0
WATER DISPOSITION			
Return to atmosphere:			
Dust suppression (evaporation)	825	187	15.1
Cooling tower (evaporation)	1,000	227	18.3
Hydrogen unit CO_2 vent(4)	445	101	8.1
Flare steam, misc. vents and losses	475	108	8.7
Retorting flue gas stacks	660	150	12.1
Shale moisturizing stacks	500	114	9.1
Total return	3,905	887	71.4
Revegetation	70	16	1.3
Spent shale moisture	1,350(5)	306	24.6
Sanitary effluent	150	34	2.7
Total disposition	5,475	1,243	100.0

---

(1) For mining, crushing and spent shale disposal.
(2) Equivalent to 8,000 acre-ft/year, or 11.1 cubic ft/sec.
(3) Water formed chemically during retorting of shale, and from fuel burned in ball heater and ball circulation system.
(4) 267 gpm of water vapor, plus 178 gpm of steam used to produce CO_2 and hydrogen.
(5) The difference between 1,350 gpm for spent shale moisture and 820 gpm of shale moisturizing makeup is supplied by use of stripped sour water and other treated recycled effluent waters.

-- The topography of the selected disposal site, as well as the sub-surface geology and hydrology.
-- The permeability of the spent shale surface embankment, as well as the leaching characteristics of the spent shale.
-- The structural integrity, erosion potential and liquefaction potential of the spent shale embankment.
-- The degree to which revegetation of the embankment surface can be achieved.
-- The feasibility of catchment dams to collect and divert surface run-off around the embankment.
-- The amount of rainfall and snowmelt at the selected disposal site, as well as the balance between precipitation and natural surface evaporation.
-- The drainage systems that might be designed for the disposal embankment.

It would be beyond the scope of this report to attempt any quantitative estimate of the pollution potential from the spent shale disposal. However, it is a serious point of concern and one which will require a great deal of study for each specific project site.

SPENT SHALE DISPOSAL

The amount of spent shale for disposal from the specific design discussed herein will be about 53,750 tons/day. When moisturized with 12% water, the spent shale sent to disposal will be about 61,000 tons/day. Over a 20-year period, this will amount to about 450 million tons. It has been estimated that disposal of that amount of spent shale would require about 800 acres covered to an average depth of 350 feet.

The disposal of spent shale is probably the major environmental factor involved in a shale oil production project. The amount of land required for the disposal site, the revegetation and reclamation of that land, and the potential for water pollution from the spent shale embankment are major problems that must be recognized and effectively dealt with.

OTHER WASTES

Shale Dust -- Shale dust from the retorting units, in the form of a wet sludge from the vent stack venturi scrubbers, will amount to about 950 tons/day. This has been included in the estimated 53,750 tons/day of spent shale for disposal. An additional 425 tons/day of raw shale dust from the dust collection systems (at the crushers and the shale storage silos) will be sent to the spent shale disposal site.

Catalysts -- Spent catalysts in the various shale oil refining units must be replaced with fresh catalysts at intervals varying from 1 to 5 years. The maximum amount of spent catalyst involved is about 850 tons/year on an annualized basis -- which is a relatively insignificant tonnage compared to the 22 million tons/year of moisturized spent shale. These spent catalysts will be sent to the spent shale embankment.

About 535 tons/year (of the 850 tons/year total) will be a material used to remove arsenic from the shale oil naphtha and gas oil. This material will contain about 107 tons/year of arsenic, of which about 29 ppm will be water-soluble. As a result, perhaps 6.2 pounds per year of water-soluble arsenic could potentially be leached from the spent shale embankment. However, the spent shale itself contains perhaps 0.11 ppm of water-soluble arsenic, which would amount to 4,840 pounds per year in the 22 million tons/year of moisturized spent shale. Thus, although the spent catalyst will contribute relatively very little potential arsenic leachate, the overall arsenic leachate potential must be carefully investigated.

Filter media -- About 215 tons per year of diatomaceous earth and 215 tons per year of spent activated carbon will be discharged from filters used in amine systems for gas treating. These materials will probably be sent to the spent shale embankment.

Coke -- The 800 tons/day of coke produced in the shale oil project will either be sold as fuel, or sent to the spent shale embankment.

Spent Caustic -- About 2 tons/day of spent sulfidic caustic will be disposed of in the spent shale.

Sludges -- Sludges from fresh water clarification and demineralization, oily water API separators, and sanitary sewage treatment will all be sent to the spent shale embankment.

In summary, these materials might all be sent to the spent shale disposal site:

	Tons/day	Tons/year
Spent shale (dry)	53,750	19,619,000
Spent shale moisture(1)	7,300	2,665,000
Raw shale dust	425	155,000
Spent catalysts	-	850
Filter media	-	430
Coke(2)	800	292,000
Spent caustic	2	730
Water treating sludges	?	?

(1) Stripped sour water, boiler blowdown, cooling tower blowdown, treated wastewater, demineralizer rinse waters, etc.

(2) Will be sold as fuel, if possible.

OTHER ENVIRONMENTAL FACTORS

The mining of 21,700,000 tons per year of oil shale (66,000 tons per day for about 328 days per year) is a very major operation. The disposal of some 22,500,000 tons per year of spent shale and other solid wastes is an equally large operation. Although it is beyond the scope of this report to do other than quantify these operations, it should be emphasized that they will cause major environmental concerns and they must be carefully dealt with.

The air emission and water balance factors have been discussed and quantified. Other environmental factors are relatively minor by comparison, but are briefly discussed in the following sub-sections.

In-plant noise will be a distinct problem, but not an insurmountable one. The mining and crushing operations, in particular, may require administrative controls in addition to engineering design controls. A limit of 60-70 dBA at the plant property line should be realistically attainable.

A number of storage tanks will be required for plant products and intermediates, as well as for chemicals, catalysts and water supply. These may total to about 1,500,000 barrels of storage.

A large emergency flare system will be needed. When flaring at maximum emergency conditions, the flame will be quite high and very noisy. However, this condition should occur only rarely. Under normal conditions, the amount of flaring should be rather nominal and it can be designed to be smokeless.

The plant will probably require access roads, a railroad spur, and a 20 to 30-inch water supply pipeline. These might require a total of 300-350 acres in addition to the land required for mining, processing and spent shale disposal.

The plant and mine will require an operating staff of about 800-1,000 people. The peak construction staff will number perhaps 2,000 and the total construction period may be 3-4 years. These operating and construction personnel will create a number of housing and other socio-economic impacts, some of which will be permanent and others will be relatively temporary.

OTHER PROCESS CONFIGURATIONS

The project described herein is based on the TOSCO II retorting system. As discussed earlier herein, there are at least three other retorting systems that might be used. As an additional variable, underground in-situ combustion might be used. These other options, if selected, might significantly change the emission and disposal factors presented in this section.

The degree to which the shale oil is refined could also vary quite widely from project to project. Again, this could significantly change the emission and disposal factors.

The recovery of hacohlite (sodium bicarbonate) and dawsonite (an aluminum-bearing mineral) along with shale oil would undoubtedly change the emission and disposal factors presented in this section.

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SECTION XII

COAL LIQUEFACTION

As discussed in an earlier section herein, coal can be gasified to produce a clean, sulfur-free gas essentially of the same quality as natural gas. The gasification of coal also produces some by-product liquid fuels, amounting to 15-20% of the total fuel values produced.

As an alternative to gasification, coal may be converted, either by pyrolysis or by dissolution in a solvent, into a range of fuels. These include clean gas, low-sulfur fuel oils or synthetic crude oil, solid char, and solvent refined coal. All of the pyrolysis and dissolution processes are broadly referred to as 'coal liquefaction' -- although that is somewhat of a misnomer since the end products include gas and solid fuels as well as liquid fuels.

Pyrolysis involves heating the coal at pressures of about 10 psig to strip out the volatile hydrocarbons, and then catalytically hydrotreating the hydrocarbon liquids to desulfurize them. Relatively large amounts of gas and solid char are produced along with the hydrocarbon liquids. Some of the gas or the char can be converted to supply the hydrogen needed for hydrotreating the liquid products. Alternatively, the char can be gasified to produce additional fuel gas product. The heat required for the pyrolysis processes may be obtained by burning some of the pyrolysis char with either oxygen or air.

The dissolution processes actually dissolve coal in a hydrogenated solvent oil at temperatures of 750-850°F and pressures of 150 to 2,500 psig. The end products (after removal and recovery of the solvent oil) include gas, oil, and char or a coking feedstock. In one case, a solid fuel is produced (solvent refined coal). The dissolution processes can be sub-classified into three categories:

1 -- Those which do not use a catalyst or a hydrogen gas supply in the dissolution reaction.

2 -- Those which do not use a catalyst, but which do use a hydrogen gas supply in the dissolution reaction.
3 -- Those which use both catalyst and a hydrogen gas supply in the dissolution reaction.

Table 15 lists most of the coal liquefaction processes currently in development in the U.S. There are three pyrolysis processes included in Table 15. Although the pyrolysis itself does not utilize either catalyst or hydrogen, the subsequent hydrotreating of the product oil does require a hydrogen supply.

The six dissolution processes in Table 15 are sub-classified (see above discussion) as to their usage of catalyst or hydrogen in the dissolution reaction. As shown in Table 15, all of the non-catalytic dissolution processes will require a subsequent hydrotreater for their oil products. The hydrotreating desulfurizes the oil and also provides a hydrogenated solvent oil. On the other hand, the catalytic processes, which require hydrogen gas for the dissolution reaction, do not need the subsequent hydrotreating of product oils. The hydrogen environment in the catalytic dissolution simultaneously causes the product oils to be hydrogenated.

In any event, all of the processes (pyrolysis and dissolution) need a hydrogen supply for one purpose or another. The hydrogen may be supplied by an external source, by conversion of some product gas, or by gasification of product char. In some cases, the coke produced from the product coking feedstock may be subsequently gasified to provide a hydrogen supply.

Table 16 presents a brief summary of the developmental status of the various coal liquefaction processes. None of the processes has yet been fully developed. Unless their funding and development are dramatically accelerated, it will probably be 4-6 years before any of these programs result in a large commercial plant.

TABLE 15 -- COAL LIQUEFACTION PROCESSES

DEVELOPER	PROCESS NAME	FUEL PRODUCTS	DISSOLUTION REQUIRES:		CATALYTIC HYDROTREATER FOR OIL
			CATALYST	H_2	
Pyrolysis Processes:					
FMC Corp.	COED (Char Oil Energy Development)	Gas, oil and char(1)	N/A	N/A	Yes
Garrett Res. & Dev.	Coal Pyrolysis	Gas, oil and char(1)	N/A	N/A	Yes
The Oil Shale Corp.	TOSCOAL	Gas, oil and char(1)	N/A	N/A	Yes
Dissolution Processes:					
Consolidation Coal Co.	CSF (Consol Synthetic Fuel)	Gas and oil(2)	No	No	Yes
Exxon Res. & Eng.(3)	Exxon Process	Gas and oil(2)	No	No	Yes
Pittsburgh and Midway Mining Co. (PAMCO)	SRC (Solvent Refined Coal)	Gas, oil and refined coal	No	Yes	Yes
Hydrocarbon Research (HRI)	H-Coal	Gas, oil and coker feed(4)	Yes	Yes	No
Bureau of Mines	Synthoil	Gas, oil and coker feed(4)	Yes	Yes	No
Gulf Res. & Dev.	CCL (Catalytic Coal Liquids)	Gas, oil and coker feed(4)	Yes	Yes	No

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(1) Char most probably gasified to supply low-Btu fuel gas

(2) Produces char which will most probably be converted to supply hydrogen for hydrotreater

(3) The technical literature indicates this process to be similar to CSF process

(4) The product coke might possibly be converted to supply hydrogen to dissolution reactor

TABLE 16 -- DEVELOPMENTAL STATUS OF COAL LIQUEFACTION

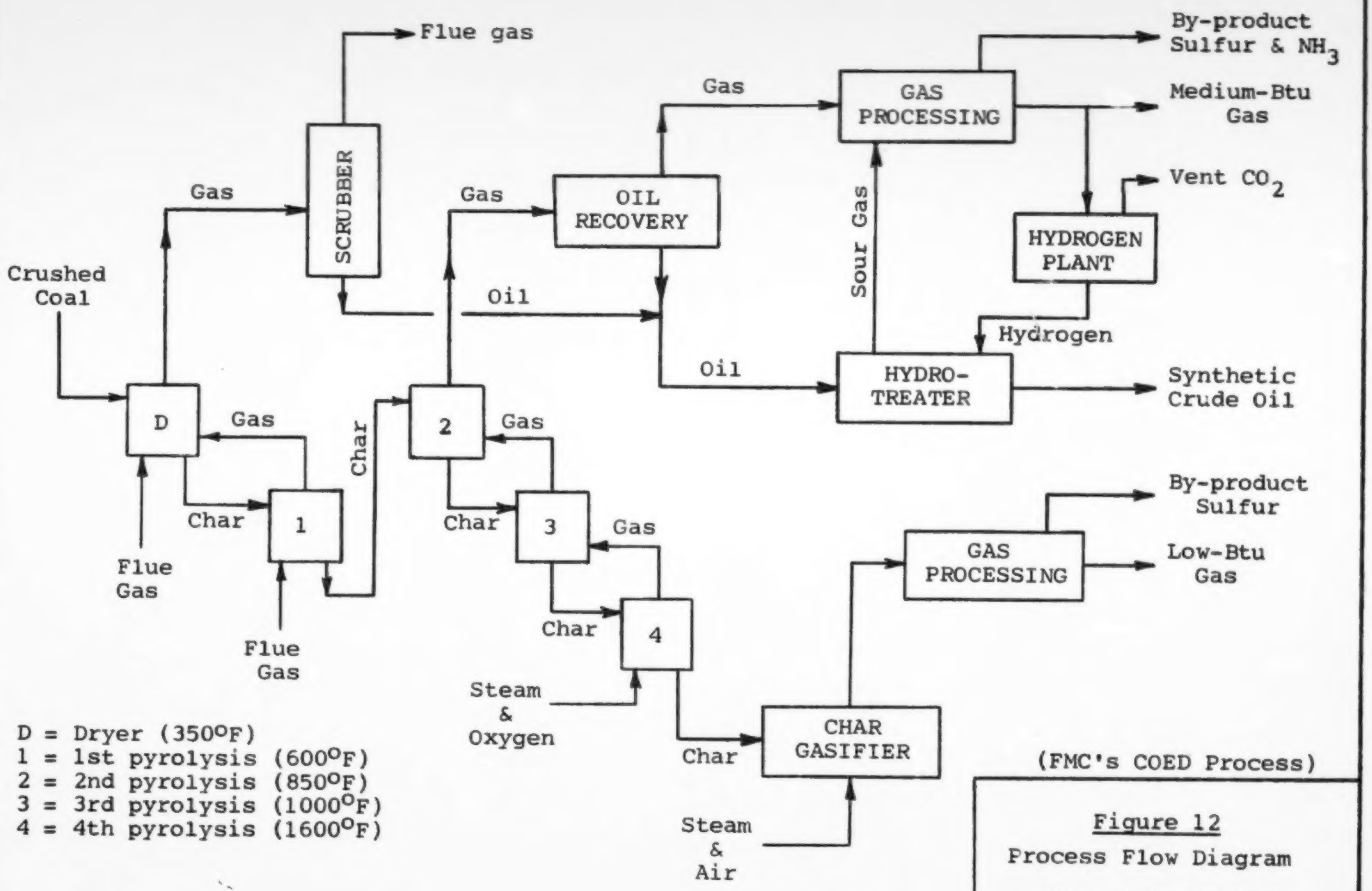
PROCESS	STATUS
COED	12 years of development by the FMC Corporation and the U.S. Office of Coal Research. A 36 ton/day pilot plant has operated since 1970.
Garrett	5 years of development. A 3.6 ton/day pilot plant is currently in operation.
TOSCOAL	Test runs have been made in a 25 ton/day pilot plant. Analagous experience has been obtained in 1,000 ton/day semi-works oil shale unit.
CSF	A 70 ton/day pilot plant was shut down after 40 months of testing, with less than 20 days of onstream operation. A detailed review of operating problems and process improvements is now underway.
Exxon	No information available.
PAMCO SRC	A 50 ton/day pilot plant is now under construction.
H-Coal	A 3 ton/day pilot plant has been operated. A 700 ton/day demonstration unit has been proposed as the next step of development.
Synthoil	A ½ ton/day pilot plant is currently in operation. Planning is underway for a 5-10 ton/day pilot plant.
Gulf CCL	Development work has been done on a small 120 lb/day unit. A 1 ton/day pilot plant is in planning.

TYPICAL PROCESS FLOW DIAGRAMS

Although some detailed designs have undoubtedly been prepared by the coal liquefaction process developers and their engineering-contractors, such designs are not readily available in the published literature. The process flow diagrams discussed herein are only conceptual and do not reflect a detailed design. However, they will serve to illustrate the basic process concepts.

Figure 12 is a flow diagram of a coal liquefaction plant based on the COED process listed in Table 15. The crushed coal feedstock is first dried with hot flue gas, and then subjected to four stages of low-pressure (i.e. 10 psig) pyrolysis. The pyrolysis temperature ranges from $600^{\circ}F$ in the first stage to about $1600^{\circ}F$ in the fourth stage. Some of the char is burned with oxygen in the fourth stage, and the hot gas from that stage provides heat for preceding stages. The hydrocarbon gases from the pyrolysis stages are cooled (or scrubbed) to produce a raw gas and a raw oil. The gas is processed for removal of H_2S and NH_3, which are converted into by-product sulfur and NH_3. The resulting product is a clean, medium-Btu gas. Some of that gas is converted to hydrogen which supplies the product oil hydrotreater. The pyrolysis char is gasified to produce a low-Btu gas.

Figure 13 is a coal liquefaction flow diagram based on the Garrett pyrolysis process listed in Table 15. The crushed coal feedstock is first carbonized at $1100^{\circ}F$ in an entrained-bed reactor. The resulting hydrocarbon gases and char are then separated. Part of the char is burned in a char heater which provides a hot char recycle to supply heat for the carbonizer. The hydrocarbon gases are cooled and scrubbed to yield a raw gas and a raw oil. Then the gas is processed for removal and recovery of by-product sulfur and NH_3. Some of the clean gas product is converted to hydrogen which supplies the product oil hydrotreater. As noted in Table 15, the product char could be converted to produce a low-Btu gas product.


Flue gas
By-product Sulfur & NH$_3$
Gas
GAS PROCESSING
Medium-Btu Gas
Gas
SCRUBBER
Gas
OIL RECOVERY
Vent CO$_2$
Sour Gas
HYDROGEN PLANT
Crushed Coal
Oil
Hydrogen
Oil
HYDRO-TREATER
Synthetic Crude Oil
D
Gas
Char
2
Gas
Char
1
Char
3
Gas
By-product Sulfur
Flue Gas
GAS PROCESSING
Low-Btu Gas
Flue Gas
Char
4
Steam & Oxygen
CHAR GASIFIER
Char
Steam & Air

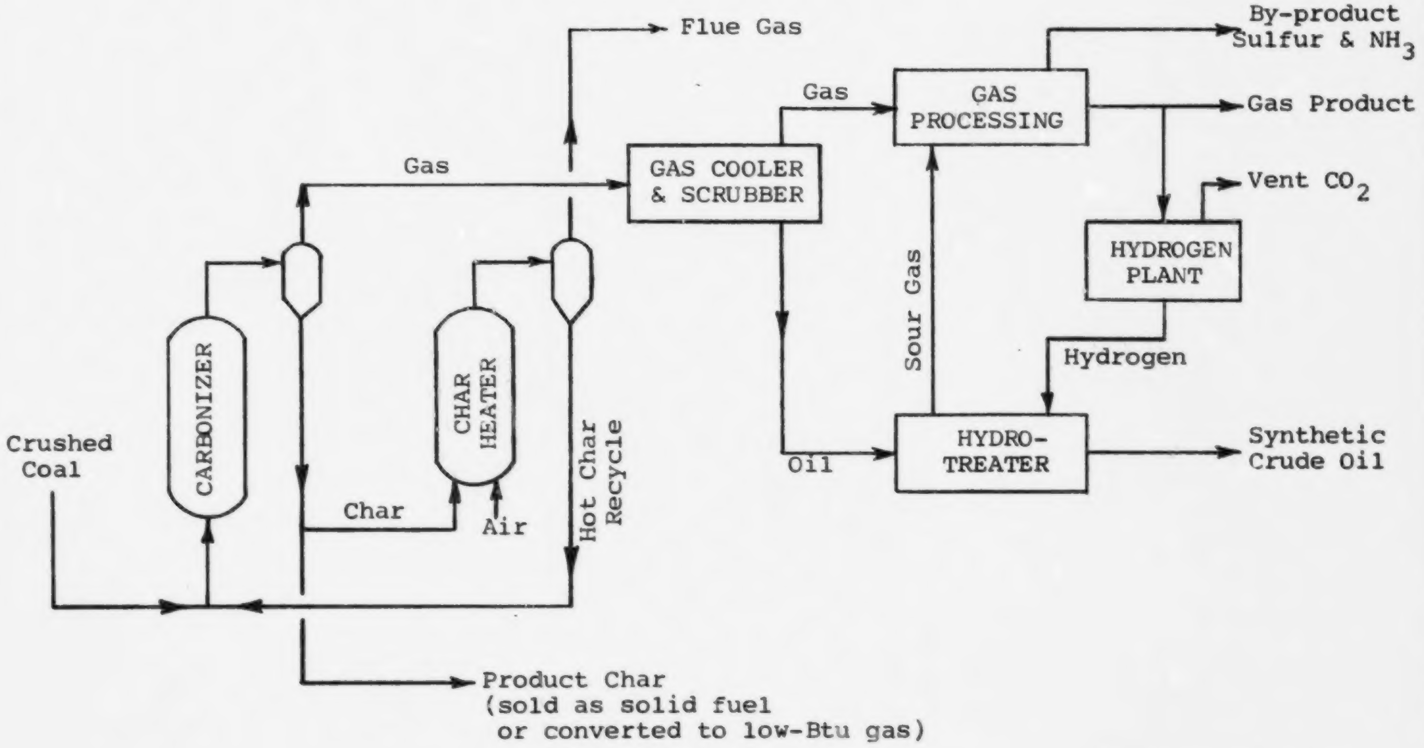
D = Dryer (350°F)
1 = 1st pyrolysis (600°F)
2 = 2nd pyrolysis (850°F)
3 = 3rd pyrolysis (1000°F)
4 = 4th pyrolysis (1600°F)

(FMC's COED Process)

Figure 12

Process Flow Diagram

COAL LIQUEFACTION
(COED PROCESS)


Flue Gas
By-product
Sulfur & NH3
Gas
GAS PROCESSING
Gas Product
Gas
GAS COOLER & SCRUBBER
Vent CO2
HYDROGEN PLANT
Sour Gas
Hydrogen
CARBONIZER
CHAR HEATER
Hot Char Recycle
Crushed Coal
Char
Air
Oil
HYDRO-TREATER
Synthetic Crude Oil
Product Char
(sold as solid fuel
or converted to low-Btu gas)

(Garrett Process)

Figure 13

Process Flow Diagram

COAL LIQUEFACTION
(GARRETT PROCESS)

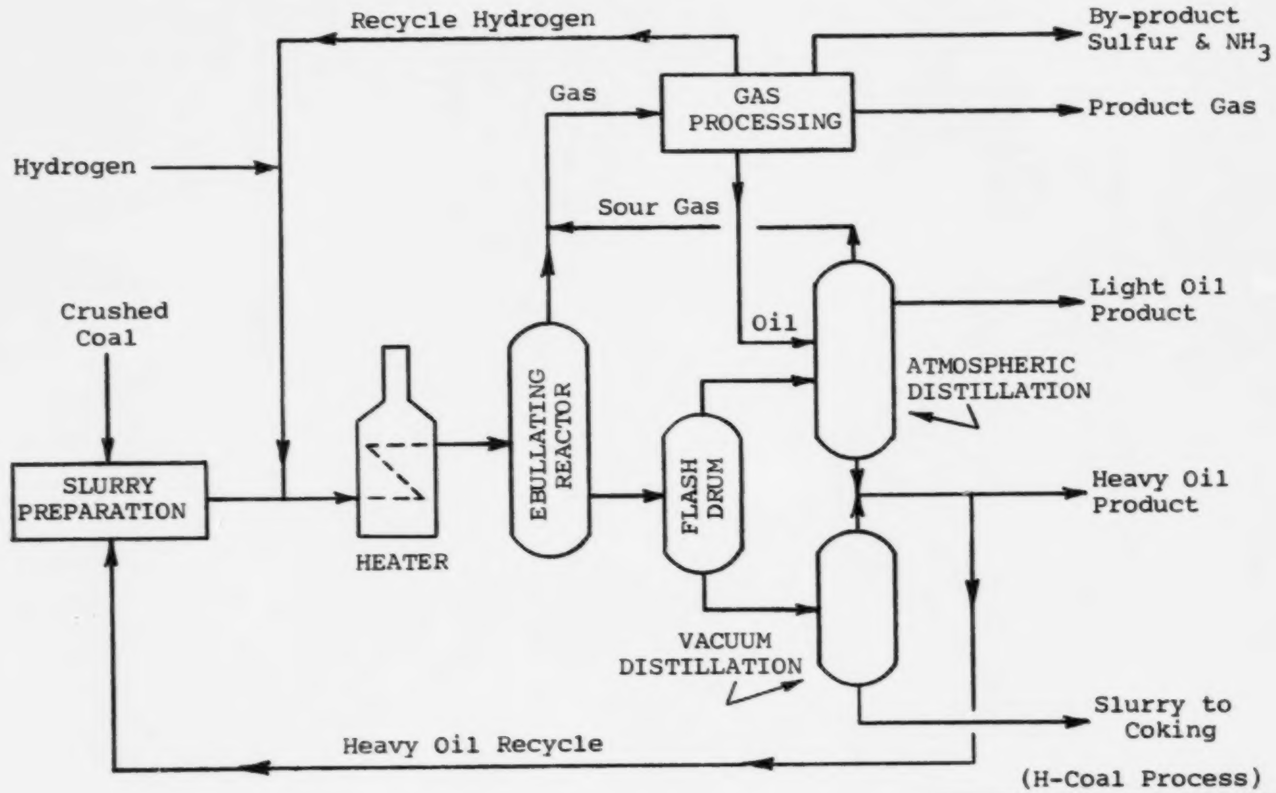
The TOSCOAL pyrolysis process, although not illustrated in this section, would utilize the same basic design as shown in Figure 11 (in Section XI on shale oil production). The TOSCOAL process would be essentially the same as the TOSCO II oil shale retorting process, except that crushed coal would be processed rather than oil shale. The resulting pyrolysis oils would require hydrotreating to produce low-sulfur fuel oils or a synthetic crude oil.

Figure 14 is a coal liquefaction plant flow diagram based on the catalytic dissolution H-Coal process listed in Table 15. The crushed coal feedstock is slurried with a recycle of heavy oil product, mixed with hydrogen gas, heated to about 850^{o}F and then reacted at about 2,500 psig. The raw gas from the reactor is processed for removal and recovery of by-product sulfur and NH_3. The gas processing also separates unused hydrogen for recycle to the dissolution reaction. The raw oil from the high-pressure reactor is 'flashed' down to low-pressure. The flashed vapors are distilled in an atmospheric pressure distillation unit, and the flashed liquid is distilled under vacuum. The distilled products are low-sulfur fuel oils, a heavy oil recycle to the coal slurry preparation, and a residuum slurry which can be fed to a coking plant. Figure 14 indicates an external source of hydrogen gas for the dissolution reaction -- but the coke produced from the residuum slurry could be gasified to provide a hydrogen supply.

These process flow descriptions typify the concepts involved in the pyrolysis and the dissolution liquefaction processes. Flow diagrams for the other processes listed in Table 15 are readily available in the technical literature (see the additional reading list at the end of this section).

OVERALL THERMAL EFFICIENCY

As best as can be determined from the available literature, the overall thermal efficiency of the coal liquefaction processes is about 65% -- which compares rather well with the 70% thermal effic-


Recycle Hydrogen
Gas
GAS PROCESSING
By-product Sulfur & NH_3
Product Gas
Hydrogen
Sour Gas
Crushed Coal
Oil
Light Oil Product
ATMOSPHERIC DISTILLATION
SLURRY PREPARATION
HEATER
EBULLATING REACTOR
FLASH DRUM
Heavy Oil Product
VACUUM DISTILLATION
Slurry to Coking
Heavy Oil Recycle

(H-Coal Process)

Figure 14

Process Flow Diagram

COAL LIQUEFACTION
(H COAL PROCESS)

iency for a Lurgi coal gasification project.

The thermal balance for a COED liquefaction plant (see Figure 12) processing 25,000 tons/day of Illinois coal would approximate the following:

	10^9 Btu/day
INPUTS:	
25,000 tons/day of coal @ 12,500 Btu/lb	625
3,750 ton/day of purchased oxygen(1)	10
190 MW of purchased electrical power(2)	46
	681
OUTPUTS:	
26,000 BSD of oil @ 5.9 x 10^6 Btu/barrel(3)	153
1,330 MM SCFD of gas @ 215 Btu/SCF	286
	439(4)

(1) This assumes a total compression requirement of 56,250 HP to cryogenically separate and provide oxygen at about 10 psig, and this amounts to 360 HP-hr per ton of oxygen. It also assumes that the compression will require about 7488 Btu/HP-hr, which is equivalent to a thermal energy efficiency of about 34%.

(2) This assumes that the electrical power generation will require 10,045 Btu/KW-hr, which is equivalent to a thermal energy efficiency of about 34%.

(3) This is equivalent to a 29 °API product oil with a heating value of 19,000 Btu/lb.

(4) This is the total net output of the liquefaction plant after supplying the plant fuel needs.

The overall thermal efficiency for the above COED process balance is 64.5%, when including the energy required to produce the purchased oxygen and electric power.

A similar thermal balance for a CSF liquefaction plant (see Table 15) processing 23,360 tons per day of a Pennsylvania coal would approximate the following:

	10^9 Btu/day
INPUTS:	
23,360 tons/day of coal @ 10,830 Btu/lb	506
3,180 tons/day of purchased oxygen(1)	15
	521
OUTPUTS:	
47,600 BSD of oil @ 6 x 10^6 Btu/barrel	285
Product fuel gas	75
Residuum oil	85
Plant fuel (heaters, boilers, power generation)	-106
	339

(1) This assumes a total compression requirement of 82,810 HP to provide oxygen at 1,000-1,200 psig, which amounts to 625 HP-hr per ton of oxygen. It also assumes that the compression will require about 7488 Btu/HP-hr, which is equivalent to a thermal energy efficiency of about 34%.

The overall thermal efficiency for the above CSF process balance is 65.1%, when including the energy required to produce the purchased oxygen. Electric power has been generated in-plant in this case, rather than being purchased.

SULFUR EMISSIONS

Lacking a detailed design for a coal liquefaction plant, it is not possible to quantify an overall material or sulfur balance. Nor is it possible to quantify stack emissions or water balance. However, generalizations and estimates can be made.

One of the major functions of a coal liquefaction plant would be to produce clean, low-sulfur fuels from a relatively high-sulfur coal -- so we can base our estimates on such coals. We can assume that the difference between the heating value of the feedstock coal and of the net plant products is consumed as plant fuel, utilized as reaction heat, and rejected as heat loss. We can further assume that 60% of the difference is plant fuel and 40% is reaction heat and heat loss. These assumptions make it possible to estimate an overall sulfur balance for given sulfur contents in the coal and the products. For example, from the COED thermal balance discussed above:

Estimated plant fuel = 0.6 (625-439)
= 111.6 x 10^9 Btu/day (18,900 BSD)

Assuming that the Illinois coal has 3.5 wt % sulfur and the product sulfur levels shown below, we obtain:

	tons/day sulfur
INPUT: 25,000 tons/day coal (3.5 wt % S)	875
OUTPUTS:	
26,000 BSD product oil (0.3 wt % S)	12
1,330 MM SCFD product gas (10 gr H_2S/100 SCF)	10
18,900 BSD plant fuel oil (0.3 wt % S)	9 *
By-product sulfur (@ 99.5% recovery)	840
Sulfur recovery tail gas	4 *
	875

* Total sulfur emissions = 13 tons/day

As another example, from the CSF thermal balance discussed above:

Plant fuel (as given) = 106 x 10^9 Btu/day
= 85 x 10^9 Btu/day residuum (12,700 BSD)
+ 21 x 10^9 Btu/day gas (98 MM SCFD)

Assuming that the Pennsylvania coal has 4.0 wt % sulfur and the product sulfur levels shown below, we obtain:

	tons/day sulfur
INPUT: 23,360 tons/day coal (4.0 wt % S)	934.4
OUTPUTS:	
47,600 BSD product oil (0.3 wt % S)	22.0
251 MMSCFD net product gas (10 gr H_2S/100 SCF)	1.8
12,700 BSD residuum fuel oil (0.7 wt % S)	15.3 *
98 MM SCFD fuel gas (10 gr H_2S/100 SCF)	0.7 *
By-product sulfur (@ 99.5% recovery)	890.1
Sulfur recovery tail gas	4.5 *
	934.4

* Total sulfur emissions = 20.5 tons/day

These estimated sulfur balances indicate that coal liquefaction plants processing 23,000-25,000 tons/day of coal having 3.5-4.0 wt % sulfur will release 13-21 tons/day of sulfur to the atmosphere. The equivalent SO_2 release would be 26-42 tons/day. These estimates are based on having sulfur recovery units (within the liquefaction plants) which are designed to achieve 99.5% recovery of process H_2S. The resulting yield of by-product sulfur will range from 840-890 tons/day.

OTHER ENVIRONMENTAL FACTORS

As a broad generalization, all of the other environmental factors involved in coal liquefaction would probably be of the same magnitude as a coal gasification plant (see Section V).

Probably the major environmental problems might center around the mining of 25,000 tons/day of coal. However, since most of our high-sulfur coal is in our Eastern and Mid-Western states, the coal would probably be mined underground rather than strip-mined on the surface. Underground mining might have less environmental impact than strip-mining.

Although a water balance cannot be estimated without a detailed design being available, we can assume that the coal liquefaction plants will require a large supply of fresh water -- but perhaps somewhat less than needed by a comparable coal gasification plant.

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SECTION XIII
GLOSSARY

Terminology	Definition
Acid gases	H_2S and CO_2
Associated gas	Raw natural gas occurring with crude oil from the same well
C_1, C_2, C_3, C_4, etc.	Hydrocarbons with 1, 2, 3, 4, etc. carbon atoms
Condensate, or natural gasoline	Liquid C_5, C_6, C_7, etc. derived from raw natural gas
Dry gas	A gas containing primarily C_1 and C_2, with very little C_3 or heavier hydrocarbons
Field gas plants, gas treating plants, gas recovery plants	Plants which process raw natural gas to recover liquids, to remove H_2S and CO_2, and to remove water
Flue gas, stack gas	Synonymous terms for the gases resulting from the combustion of a fuel
Heavy hydrocarbons	Higher-boiling, higher-density hydrocarbons with about 7 or more carbon atoms
Heavy oils	Fuel oil, heavy distillate oil, heavy furnace oil, No. 6 oil, bunker oil, resid. Hydrocarbon mixtures of from C_{16} to C_{20+} and boiling from 650 to 1000+°F, derived from crude oil
High-Btu gas	Fuel gas with higher heating value of about 1000 Btu/SCF or more
Higher heating value, gross heating value	The total heat released when a fuel is burned
Hydrocarbon	Chemical molecule composed of hydrogen and carbon atoms
Light gasoline, light naphtha	Liquid C_5, C_6, C_7, C_8 derived from crude oil. About the same as condensate from raw natural gas
Light hydrocarbons, light ends	Low-boiling, low-density hydrocarbons with from 1 to 6 carbon atoms
Low-Btu gas	Fuel gas with higher heating value of 350-450 Btu/SCF or less
Lower heating value, net heating value	The effective usable heat released when a fuel is burned (after some gross heat release is used in vaporizing the combustion product water)

Terminology	Definition
Middle distillates	Kerosene, diesel oil, jet fuel, light furnace oil. Hydrocarbon mixtures of from C_{10} to C_{15} and boiling in the range of 350 to 650°F, derived from crude oil
Naphtha, gasoline, full range naphtha	A hydrocarbon liquid mixture of C_5 to C_{11} and boiling in the range from 100 to 400°F
% excess air	The percentage of combustion air supplied to a burning fuel, over and above that required to combine with the fuel hydrogen and carbon
% O_2 in flue gas	The % of oxygen in the flue gas, as a result of using excess combustion air
Sour gas	A gas containing H_2S
Sweet gas	A gas containing little or no H_2S
Thermal efficiency	The percentage of the total heating value input (including fuels) to a plant that is recovered as product and byproduct heating value or equivalent energy
Virgin naphtha, virgin oils	Naphtha and oils derived from atmospheric and vacuum distillation of crude oil
Wet gas	A gas containing significant amounts of C_3 or heavier hydrocarbons

Abbreviations and Units	Definition
API	American Petroleum Institute
Btu	British thermal unit, the amount of heat required to raise the temperature of 1 pound of water by 1°F
CO, CO_2	Carbon monoxide and carbon dioxide gases
DFG	Dry flue gas
gpm	Gallons per minute
H_2, H_2O	Hydrogen and water
H_2S	Hydrogen sulfide gas
HHV	Higher heating value
LHV	Lower heating value
LNG	Liquefied natural gas, mostly methane
LPG	Liquefied petroleum gases, usually C_3 and C_4

Abbreviations and Units	Definition
MM, or 10^6	Synonymous terms for the number 1 million (or 1,000,000)
NGL	Natural gas liquids, a collective name for C_3 LPG, C_4 LPG, and C_5-C_8 condensate
O_2, N_2	Oxygen and nitrogen gases
ppmv	Parts per million by volume
ppmw	Parts per million by weight
ROW	Right-of-way
S	Sulfur
SCF	A standard cubic foot of gas, measured at atmospheric pressure and 60°F
SCFD	Standard cubic feet per day
SNG	Substitute or synthetic natural gas
SPG	Substitute or synthetic pipeline gas. Synonymous with SNG
WFG	Wet flue gas

Equivalents

1 lb of S	equals	2 lbs. of SO_2
1 lb of H_2S	equals	11.15 SCF of H_2S
1 lb of SO_2	equals	5.92 SCF of SO_2
1 lb of NO_x	equals	8.24 SCF of NO_x
1 lb of O_2	equals	11.84 SCF of O_2
1 lb of N_2	equals	13.54 SCF of N_2
1 lb of H_2O	equals	21.06 SCF of H_2O
1 lb of CO_2	equals	8.61 SCF of CO_2
1 lb of flue gas	equals	ca. 13.07 SCF of flue gas
1 lb of natural gas	equals	ca. 23.69 SCF of natural gas
1 gpm of water flow	equals	500 lbs/hr of water flow
1 day	equals	1,440 minutes
1 ton	equals	2,000 pounds (a short ton)
1 barrel	equals	42 gallons